PROCEEDINGS

OF

THE ELEVENTH ANNUAL MEETING

OF

THE CANADIAN MINERAL PROCESSORS

HELD AT

CANADA CENTRE FOR MINERAL AND ENERGY TECHNOLOGY
DEPARTMENT OF ENERGY, MINES AND RESOURCES
OTTAWA

JANUARY 23, 24, 25, 1979

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HELD UNDER THE ASPICES

OF THE

CANADA CENTRE FOR MINERAL AND ENERGY TECHNOLOGY
OTTAWA, ONTARIO

JANUARY 23, 24, 25, 1979

COMPLIED BY

R.W. BRUCE, P. ENG.
SECRETARY, CMP.

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PREFACE

This publication has been prepared to provide a permanent record of the Proceedings of the Eleventh Annual Meeting of the Canadian Mineral Processors which was held at Camsell Hall, Department of Energy, Mines and Resources, Ottawa, Ontario, Canada, January 23, 24, 25, 1979.

The papers that were presented have been incorporated into this volume with the pertinent general discussion period which followed. It has not been possible to record verbatim all of the comments but every effort has been made to express the ideas as accurately as possible from tape transcriptions.

The technical program was prepared by the Program Committee which was drawn from members of the Canadian Mineral Processors under the capable chairmanship of Mr. P.G. Claridge.

The next annual meeting of the Canadian Mineral Processors will be held on January 22, 23, 24, 1980, in Ottawa. Departmental support will come through the Mineral Sciences Laboratories of CANMET.

ACKNOWLEDGEMENTS

The support and assistance of CANMET staff at the meeting is acknowledged and appreciated.

The tapes were transcribed and edited by Mr. H.H. McCreedy.

The Canadian Mineral Processors are indebted to Dr. D.F. Coates, Director-General, CANMET, and Mr. W.A. Gow, Chief of the Mineral Sciences Laboratories, whose most welcomed support helped to ensure a successful meeting.

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ATTENDANCE - ELEVENTH ANNUAL MEETING
CANADIAN MINERAL PROCESSORS
January 23, 24, 25, 1979

Name	Company	Region
M.J. Allan	Afton Mines Ltd.	B.C.
T.R. Armstrong	Ontario Hydro	Central Ontario
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D. Bridges	Mattabi Mines Ltd.	Red Lake
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R. Fish	Canadian Mining Journal	Montreal
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R.H.D. Halupka	Texasgulf Canada Ltd.	Porcupine
V.F. Harrison	Retired	Ottawa
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G. Holmes	Bell Copper	B.C.
T.M. Hill	Texasgulf Canada Ltd.	Porcupine

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N.A. Lett	Adams Mine	Central Ontario
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H. Lueske	Selco Mining Corp.	Red Lake
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D. MacGillivray	Wabush Mines	N.E. Quebec
W.A. MacKenzie	Syncrude	Alberta
A. Mai	Iron Ore Company of Canada	N.E. Quebec
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A.L. Mular	University of British Columbia	B.C.

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C.P. Patel	Falconbridge Copper Ltd.	Red Lake
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F.G.I. Pickard	Falconbridge Nickel Mines	Toronto
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M.J. Smith	Canadian Bechtel Limited	Toronto
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K.G. Stowe	Mines Lac Mattagami Ltd.	N.W. Quebec
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Name	Company	Region
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J. Wallace	Sidbec-Normines	N.E. Quebec
M.G. Watling	Hudson Bay Mining & Smelting	Man. - Sask.
C. Webber	Sherritt Gordon Mines	Man. - Sask.
A.J. Williams	Kilborn Ltd.	Toronto
D.R. Whittall	Cominco Ltd.	Yukon - N.W.T.
K. Wood	Selco Mining	N.W. Quebec
D. Wood	Brunswick Mining	Atlantic
W.T. Yen	Salman Mineral Research Ltd.	Montreal
K.A. Young	Canadian Bechtel Limited	Toronto

PROCEEDINGS OF THE ELEVENTH ANNUAL MEETING OF THE CANADIAN MINERAL PROCESSORS

Chairman - D.M. Doyle

Chairman's Opening Remarks

Good morning ladies, gentlemen, authors, distinguished guests, welcome to the Eleventh Annual Meeting of the Canadian Mineral Processors.

My name is Don Doyle. I have had the pleasure of serving as your 1979 chairman, but my term of office expires on Thursday of this week, at which time your program chairman Phil Claridge will assume this position.

I know I speak for the regular CMP attendees when I say I am pleased to see so many familiar friendly faces in the audience. I would like in particular to address the following comments to those unfamiliar faces who have travelled from the far corners of Canada to participate in this Ottawa meeting and who may be wondering what CMP is all about and wondering what is in store for them.

The members of Canadian Mineral Processors are the operators, consultants and academics directly involved in the Canadian Mineral Industry. From a nucleus of gold operators we have expanded to now represent the gold, base metal, iron ore, potash, uranium and tar sands operations across Canada.

A good number of CMP members have spent considerable time pondering the future course of this proud organization. You will be given an opportunity at the business meeting to vote on a motion which could significantly affect our future course of direction.

I would like to introduce some CMP members who have had a significant impact on our organizaiton and on the industry we represent. Firstly, I would ask that the 1978 Regional Representatives and Executive please rise and remain standing. I now ask that all former representatives and executives please rise and remain standing. We are indebted to the staff here at CANMET for their contributions to our orgainzation, and I ask that they now rise. The technical expertise which flows through this organization is exemplified in the technical papers presented at the regional and annual meetings and I ask that all authors please rise. I hope that I speak for all these gentlemen when I say that our participation in CMP activities has been a highlight in our technical careers. Welcome to our annual meeting. Gentlemen please be seated. I hope that we all make an extra effort to further our acquaintances and expand our technical experience during the next few days.

I have the pleasure of introducing a man of extraordinary capabilities, in the field of energy, mining, mineral processing extractive metallurgy, and public speaking. Dr. Drolet has received degrees at Laval, Columbia and McGill. He has occupied various positions from miner to Executive Director of Quebec

Cartier Mining Company, from mining engineer of the Quebec Department of Mines to Assistant Deputy Minister of Energy, Mines & Resources. Would you please join me in welcoming the Senior Assistant Deputy Minister for International Minerals - The 71st President of the Canadian Institute of Mining and Metallurgy, Doctor Jean-Paul Drolet.

THE SUPPLY EQUATION:
RESOURCE BASE & PUBLIC PERCEPTION

by

Jean - Paul Drolet
President

Canadian Institute of Mining and Metallurgy

It is the second time that I have the pleasure of addressing the Canadian Mineral Processors, although I have also had in former years the opportunity to participate actively in some of your meetings. I draw this distinction because I consider quite differently the task of presenting you with the classical words of welcome, and the responsibility of the president of the Canadian Institute of Mining and Metallurgy to meet and exchange views with the members of groups concerned with mineral development. It is therefore with pleasure that I have received and accepted your invitation.

The Institute has grown tremendously in recent years. It has grown in membership which now reaches 11000 and also in its professional activities through the technical divisions and societies. I am aware that the mineral processors are presently considering the possibility of an affiliation to the CIM as a section of the Metallurgical Society, which constitutes the most important body of our Institute. I want to assure the officers and members of your group that the CIM is convinced that such initiative would be of great benefit to the mineral processors and would also bring an added prestige to the Canadian Institute of Mining and Metallurgy.

The last time we met, I spoke about mineral exploration and the role of research and development in the application of new processing technologies and how they can affect a large number of already producing mines, as well as the deposits in the immense pool of presently uneconomic resources. The choice of a theme that can offer more than a passing interest to a group of specialists such as yours is a challenge. A pile of white sheets of paper on one's working table can easily become a frightening object when one tries to identify and develop a subject that could have some bearing on the professionnal activities of those responsible for the sound management and rational development of our mineral domain.

I decided to proceed with a simple process of elimination which is very familiar to mineral processors in winning the earth's resources, and I then elected not to deal with a long series of issues whose discussion would fit better the agendas of other meetings with representatives of the mineral industry.

I have therefore chosen to say nothing about taxation and the still unsettled controversy over the problem of sharing mining revenues, nothing about the intrusion of governments in mineral ventures or about the role of the Canadian Export Dev. Corp. in assisting foreign mining developments, nothing about the North and land use and the recent announcement concerning areas of the North that are being considered for withdrawal from resource development, nor will I say anything about health and safety, about the environment and some of its regulatory overkill by government agencies or concerned citizens. I will not even mention uranium and the legislation about foreign ownership, nothing about FIRA, nothing about the shortage of skilled labour and the role and responsibilities of labour unions, less than nothing about asbestos in Quebec, about potash in Saskatchewan, about the role of CDC and Petro-Canada and a whole bag of issues, political or otherwise, which also includes GATT, MTN, UNCTAD, Common Fund, Canada-EEC, Canada-Japan, the law of the Sea -- and, while I am at it, why not throw in the next federal election, the referendum in Quebec and to embrace them all, the Unity of Canada. To be frank, I will avoid them all -- the main reason being my lack of competence to deal with most of these issues in the time which has been allowed to me for presenting the opening address to this august assembly.

It is often interesting to reminisce and talk about the "good old days". Those were the days when almost everything related to mining -- from prospecting and grub-staking to the production of healthy profits for the producer and adequate dividends for the shareholders -- was exhilarating, fascinating and deeply satisfying to those engaged in it. In a way, it was the kind of activity that created legends -- the kind of legends that are still vividly remembered by those who were in mining regions when significant mineral discoveries were made. Nobody was standing in the path of those with imagination and talent who knew that, in a country such as Canada, mining was to become a most important contributor to the prosperity of its people. The existence of governments and their agencies was known and recognized, but their presence was hardly felt. Above all, conservationists and their cousins the environmentalists had not yet been invented.

The older we get, however, the more selective our memory becomes, and the better those good old days look. But we also know that there is more work to be done and since nostalgia seldom accomplished anything let us consider one of these issues as it relates to the progress of our industry.

One aspect of the future that I would like to dwell on is the question of *growth in mineral production*, given the far-reaching concerns about *resource conservation*. I do not intend to get involved in arguments about wildflowers versus tailing ponds; that has to do with environmental conservation which is another subject. Nor do I want to paint a battle scene of growth in production versus resource conservation. I know it is natural to think of these issues as polar opposites. Looking for opposites is an appealing way of organizing the world in our minds, of simplifying it. Love versus hate, north versus south, nature versus technology, liberals versus conservatives, are subjects that easily provoke the most sparkling arguments. You can choose your good guys and your bad guys, and the arguments will follow from there.

In considering growth and resource conservation, I believe we should try to go beyond such "black-and-white" thinking. We know from our bedside novels or from our favorite TV serials that love and hate are often hard to distinguish from one another; and I suggest that growth and conservation may be equally intertwined. However, before I elaborate on that thought, let me say a few words about the concept of conservation.

Many people confuse *conservation with preservation.* In the case of minerals, it is common to hear that perhaps we should not extract and export so many mineral products because if we keep consuming and exporting them at the current rate there will be none left for our children or, at least, for the generations that will follow.

It is obvious that we are dealing here with a wish for preservation, which, in this case, we might as well call it what it is, namely "hoarding". But it does not take long to figure out that you can help a child more by teaching him how to catch his own fish than to present him with a package of frozen fish to be kept on ice for his old age. It is obvious that hoarding mineral deposits for posterity -- which is likely to be wealthier than we are -- would be a sterile and unimaginative approach. Especially so because it is rooted in the assumption that nothing will change about future consumption except its level and that the same minerals will continue to be mined in the same old ways and for the same end uses that we presently know.

We never have been and probably never will be very good at foreseeing how the ways of society will change in the future. For example, imagine anyone living before the industrial age trying to conceive a society as large and as energy-hungry as the one we live in today. If he thought like a true preservationist, he would have immediately rejected the notion that so much energy per capita could possibly be available because there would not be enough room for all the horses required to provide it, let alone enough grassland to feed them. Above all, everyone would be up to his knees in horse pollution.

Of course, a great deal is made of the fact that mineral resources are non-renewable, unlike, for instance, timber. The argument is commonly made that, since we cannot grow new mineral deposits, we must face the fact that they are in limited supply.

But I cannot see that the non-renewable nature of minerals is relevant to mineral supply over the long term. Many deposits that we are now mining would not have been considered as sources of supply some decades ago because, at that time, they could not have been mined and processed economically. We all know that the average metal grade of ore mined has decreased steadily as a result of technological improvements, without increasing real costs.

Among the various trends that may influence the growth patterns of mineral supply, it seems to me that changes in *user technology* can have the most far-reaching effects. A couple of examples could illustrate this. First, mankind at one time made extensive use of flint tools and arrowheads. The fact that we no longer do so has nothing to do with the limits of flint resources, or the costs of producing more flint. We simply grew out of the flint era. Second, peat, which is produced mostly for horticultural purposes, was once used as a prominent fuel in many countries of the world, but before we exhausted all peat resources, we progressed in time to avert a "great peat crisis".

Although we are now in what is popularly termed an "oil crisis", this refers to supply problems of a man-made nature. In any case, within half a century we will most likely have progressed beyond the need to burn hydrocarbons as our main source of energy.

Thus, an item cannot only *become* a mineral resource as a result of our requirements as we perceive them, but some resources can *cease to be* resources because we have found something better. What then is the origin of our fears about the future availability of minerals? In looking ahead, we generally do four things:

1. *We extrapolate our demand some decades into the future.*
2. *We look at the types of mineral sources presently known and at hand, and plot a plausible rate of production from these sources over the same period.*
3. *We compare these two curves and are struck by the supply gap, which widens frighteningly in the future.*
4. *We panic because we are afraid of the dark (as if the future had ever been well lit before!).*

In the past we have led ourselves to panic like this, and I am sure we will continue to do so in spite of the fact that our demand extrapolations are extremely unimaginative and that there are more mineral sources on this globe and particularly in our country than those we have already identified. If you, the mineral processors did not think it was so, you would not be looking for new and improved technologies in the recovery of these minerals.

So much for mineral conservation as viewed from the supply end. *Let us now turn to future demand.* What do we really know about it? The basic technique of demand forecasting is to observe the past and look for clues to the future, assuming that the future will be like the past. We may estimate population growth and forecast future populations. We may then estimate what such future populations will consume, as GNP per head or in terms of categories of consumer goods. From that we can forecast how much of each mineral commodity will be required.

Now, I would like to illustrate what may go wrong with such technique of forecasting. You may remember the "Paley Commission" report published in 1952, entitled "Resources for Freedom". A study was made to compare the forecasts of 25 years ago with actual events. It turned out that the growth of GNP (in real terms) over these 25 years had been substantially underestimated, while the requirements for most mineral commodities had been overestimated. This is surprising and at first somewhat baffling, but the explanation seems to be that in 1952 it was not foreseen that advances in the *efficiency of use* would reduce mineral requirements to the extent that actually took place.

For instance, we now have handy little electronic pocket calculators that replaced bulky desk calculators. They produce answers infinitely faster, although both the material input and the nergy use of these new marvels is a great deal smaller than that of their predecessors. That was the kind of growth that was not adequately foreseen. Technology forecasting is not any easier today.

We are still making this type of error: we are not adequately aware of where the changes in user technology are taking us. Surely, there will continue to be growth, but not simply growth in quantity, rather growth in quality, in versatility, and in adaptability which makes growth in other forms much easier. Think, for example of the capability that has been developed to transmit as many signals optically through a glass fibre as are presently transmitted through a copper wire of the size of a person's arm. When your friendly telephone company starts to apply this new technology universally, growth in copper demand will no doubt diminish (unless a lot of copper turns out to be needed for equipment in the challenging field of tapping those new optic-fibre telephone lines which might provide fresh avenues for growth!).

So far, I have tried to make the point that growth should not be thought of simply as "more of the same", while I have said little about conservation, except that it is different from preservation. *So, what is conservation with respect to minerals?* I believe it might refer to any efforts to get the most out of our resources. There are a number of ways to accomplish this:

- *More efficient recovery* will yield more mineral products from the same ore deposits.
- *Recycling* will stretch supplies.
- *Substitution* in the consumption of minerals can take the pressure off the scarcer minerals.
- *Efficient design* is another way; goods can be made more durable and easier to maintain and repair; they can be designed with a more frugal use of materials (think of miniaturization) and in such a way that recycling will be easier and cheaper.

It is in the cards that material conservation measures will apply more and more. In fact, automobiles have become the single largest source of scrap steel. Most of the nonferrous metals -- zinc, aluminum, copper -- contained in junk cars ended up in land fi'l not many years ago. No longer.

I could go on and talk about the effects of substitution and more efficient design on the demand patterns for primary mineral commodities, but there is no need to do so.

Therefore, I suggest that, when we look some distance ahead, we should be thinking of *growth hand-in-glove with conservation and not growth versus conservation,* at least when we refer to the conservation of minerals. However, economic growth, even if it encompasses more efficient and sophisticated uses of materials, still requires that we mine minerals. Mining means digging, and digging and subsequent ore treatments may conflict with the conservation of things that tend to be thrown into the mixed pot called "environment". We in the minerals business are only one couple of the dance floor of society. When it becomes crowded, it is necessary to pay more attention to the others. We do not want to step on their toes, elbow them in the ribs, or jostle them out of the way. The intelligent thing to do for us is to realize that, when there is crowding, all of us should stifle the urge to dance the Charleston.

The scientists among us should welcome the challenge to conceive and develop new methods that will provide satisfactory solutions. Recent advances in the art of recovering and refining metals show that many health and environmental problems can be reduced by shifting gears technologically, while overall costs may be lowered in the bargain. While we must try to resolve conflicts in value judgments on a case-by-case basis, we must also realize that many of these conflicts are a manifestation of the "the bigger, the better" syndrome, of growth for growth's sake. Such conflicts will inevitably multiply and reach more disruptive proportions unless we turn more toward the qualitative type of growth.

No society is alive that is not growing in some way. Growth is life. And, as a relatively young society, grow we must. But we have to learn to be a great deal more versatile in adapting our forms of growth to our new aspirations in a changing world.

In these particular times of political effervescence, may I conclude by suggesting that instead of having liberal growth, we should aim -- in line with our theme -- for growth that is progressively conservative; and I suppose that, in order to cover the whole political front, someone could add -- both democratic and socially creditable. In any case, as far as minerals and energy are concerned, "growth hand-in-glove with conservation" would be a good motto for Canada to adopt in the years ahead.

There are no simple solutions -- only intelligent choices. The responsibility of ensuring that the proper choices are made belongs also to the mineral processors of our resources.

Once again, I thank you sincerely for your invitation which I accepted with honour because I consider addressing the processors of Canadian minerals a unique opportunity.

Ottawa, January 11, 1979.

DESIGN AND OPERATION
OF
A FUTURE MILL

by

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INTRODUCTION

Mineral processing is undergoing growth and change at a rapid rate.

The modern flotation process was discovered only in the mid-twenties, and uranium leaching in the mid-fifties. The growth of individual unit equipment size is also a recent development and is perhaps most dramatically illustrated by the increase in size and connected horsepower in grinding mills. It is also apparent in much of the equipment used in mining operations.

These technical and mechanical developments have influenced the design of plants but not to the expected degree.

Many metallurgists believed until recently that grinding an ore in the absence of steel results in a loss of selectivity in the flotation. However results reported by Thornton (1), Fahlstrom (2), Canmet (3 & 4) have shown on the contrary better selectivity and improved recovery at a lower operating cost. According to the Aitik (2) plant results, the net difference in cost plus tailing value in favour of the two-stage autogeneous grinding circuit is in the order of 40 cents per ton of ore milled. Even a greater difference was indicated by the results obtained by Canmet on a sample supplied by Mattagami Lake Mines Ltd.

Yet how many plants are taking advantage of these findings and how many plants are being converted. How much more improvement of the results would have been obtained if these tests had incorporated optimization of classification efficiency? By nature the mining industry and the mineral processors are conservative, and tend to emphasize the possible penalties for an unsuccessful innovation out of proportion to the potential rewards even for a significant improvement.

We are still satisfied to break rock in essentially the same manner that was used thousands of years ago by pounding, rubbing and squeezing, and we still waste much energy and metal in the process.

We have succeeded only to replace human energy by mechanical energy.

Most processing plants are treating ores of variable characteristics including hardness, amount and type of associated minerals, etc., and yet many are designed to produce a finer grind on harder ore and coarser grind on the softer ores, contrary to rational economics; or without proper surge capacity between processing stages sensitive to variations; or without proper optimization of the various processing stages, resulting in voluminous circulating loads and wastage of both energy and equipment.

Devoted operators who could "read flotation" and "listen to grinding mills" used to be plentiful, but it is evident that this devotion has been on the decline in recent years, making instrumentation a necessity.

A future concentrator must be conceived to increase processing efficiency through:

1. A reduced number of processing steps from mine size to concentration size.

2. An efficient classification system, that permits the desired liberation of the valuable minerals with minimum production of slimes.

3. A selection of equipment that will permit the best metallurgy at the best cost including recovery losses.

4) A simplified flowsheet and layout.

5) An instrumentation permitting total management.

GENERAL

The mine is often concerned only with production at minimum cost of a required tonnage of ore of a specific minimum grade for the concentrator. All too frequently there is a lack of appreciation of the influence that drilling and blasting techniques, (which give rise to overbreakage and excessive production of primary slimes),or "run of mine" ore characteristics (such as hardness or mineral association) may have on plant operation, process control and metallurgical performance. For these reasons, communication and cooperation between geologists, mine planning engineers and metallurgists are necessary during testing and flowsheet development to permit the design of a plant reflecting a correct interpretation of the metallurgical process requirements.

TESTING AND FLOWSHEET DEVELOPMENT

The basis for the design of a concentrator is the thorough testing of representative ore samples in the laboratory and pilot plant.

Never trust a geologist for the selection of representative samples unless he is aware of and acquainted with the concentration problems and accepts to keep you informed of the mine developments.

Too often the samples tested do not represent the ore body or the variability of the ore. It is not unusual that good samples are not available, (even when project management is aware of the importance of testing representative samples), because the type of deposit may not permit initially the opening of many levels or areas of the mine to get these samples. However, bulk samples from the first sections of the ore body to be developed, plus core samples from other less accessible levels or zones, should be adequate normally for process testing and flowsheet development.

The testwork itself must be performed carefully, thoroughly and under realistic conditions. It must cover all aspects of the process including the economic effects of various feed preparation techniques such as crushing, grinding, classification efficiency, etc., and not only concentration techniques.

The testing should not only demonstrate a way to treat the ore but also produce a detailed evaluation of the optimum process requirements at every phase of the operation. A process flowsheet is an integrated system and the performance or efficiency of each machine or operation is influenced by the process and equipment that precede it or even follow it where middling recycling is practised.

On the other hand every one will agree that the correct technical and economic interpretation of the test data is important but very often difficult, and must be based on mature judgment and experience. Nevertheless, it is still an interpretation, and the resulting flowsheet may not be the ideal interpretation of the metallurgical requirements. With the present world inflationary trends, optimization of operations is especially important to maximize profits and to maintain our mining industry in a financially sound position. Along the same trend the management and finance people might be induced to minimize capital cost through unrealistic economies like buying the biggest at the lowest cost, instead of the best at the best cost; or by eliminating surge capacity between various processing stages. The metallurgists must be confident enough to resist to these superior pressures. Capital saving should not be made at the expense of mechanical performance or metallurgical efficiency. Compromising cost savings should be viewed with great caution, and thoroughly analyzed with practised discernment. Many bright promises of cost savings by this approach have proved to be illusory and all too often disastrous.

PROCESS STEPS IN SIZE REDUCTION

Except for large open pit mines furnished with very large loading and transportation equipment, most mines require a primary crushing stage to reduce the mined ore to a convenient top size for handling and transportation to the concentrating plant. In most small to medium size mines, this primary crushing is performed underground or at the mine.

But the secondary crushing plant, is usually located on surface, in a separate building. This set up is becoming optional in many cases and is tending to disappear as such. You will see more and more, secondary crushers set in open circuit directly ahead of the primary grinding mill as part of the grinding plant, and running continuously. The savings realized by this set-up in building, conveyors, screens and operating manpower are such that you can afford the installation of standby crushers for abrasive ore.

This tendency is not surprising to those who have recognized that crushing plant operations, on account of their high manpower for operation and maintenance, are three to four times more expensive than grinding plant operations in terms of unit of work done according to F.C. Bond formula calculation.

TABLE I

Crushing & Grinding - Direct Costs (1977)

(10,000 tons per day - Medium Hard Ore)

	$/Ton	F(µ)	P(µ)	W(kWh/T)	$/kWh
Crushing	0.24	178,000	12,700	0.65	0.37
Grinding	0.95	12,700	100	9.12	0.104

When the grinding is performed in autogeneous or semi-autogeneous mills, the secondary crushing plant is eliminated and the grinding mills are fed directly from primary crusher or mine size. This grinding method is becoming more and more accepted on various types of ore as it shows a lower capital and operating cost and very often an improved mineral liberation, and therefore better energy utilization.

The autogeneous type mills have also the advantage of being built in larger size units and these very large units offer real opportunities for advances in plant design, reduction in the number of plant lines for a given capacity and substantial savings both in capital and operating costs.

This does not necessarily mean that the minimum number of plant lines is the ideal overall plant design, but that a new plant design should be based on the concept of using the largest equipment available which is suitable. Any departure from this concept should be based on a clear demonstration of the advantages of smaller units.

The rapid increase in grinding mill sizes during recent years has not brought to light any upward limit in presently accepted grinding theory and has increased the possibility of feeding run of mine ore without the benefit of a primary crusher. Limitations of mill sizes appear to be mostly mechanical on the part of the builders and electrical on the part of the users.

Grinding mills are not employed to reduce the size of the ore "per se" but to produce effective liberation of the valuable minerals and the production of a minimum of slimes. A major portion of the recovery losses in concentration arises from poor liberation and overgrinding. These conditions are the result of wrong selection and operation of grinding mills coupled with inefficient classification systems. The grinding operation prepares the feed for the subsequent concentration operation and that feed should have such a structure quality as to produce the best metallurgy at the best cost including recovery losses. Grinding is probably the most important, usually the most costly, often the most difficult and possibly the most neglected operation in mineral processing; the real value of energy wastage due to overgrinding is not yet fully appreciated.

The industry accepts easily to spend millions of dollars to increase the grinding capacity through the addition of extra grinding units but resists the spending of thousands of dollars to improve the classification system and raise the efficiency of existing units to arrive at the same results.

For example at East Sullivan Mines Ltd., in the early fifties, a series of tests was conducted in the grinding circuit to see what could be done to increase the capacity. The addition of an extra stage of classification, using cyclones, improved the overall classification efficiency and increased the regrind mills loading. The results were amazing:

A capacity increase of 10%.

A recovery increase of 1%.

A concentrate grade increase of 1%.

The last two gains were unexpected, but due mainly to reduced overgrind.

There are two factors contributing to liberation efficiency: the retention time in the grinding mill (which is related to mill loading) and the classification efficiency.

MILL LOADING

The working capacity of a grinding mill is expressed by its power consumption. The mill feed tonnage then should be a function of the mill power and of the power per ton required at that grinding stage.

Numerous observations in different types of plants have been conducted for the development of the following "rule of thumb" formula:

$$\text{Mill Feed (in TPH)} \simeq \frac{3\ \text{kW(Mill)}}{\text{kWh/ton}}$$

i.e. the total mill feed in tons per hour equals approximately three times the mill power divided by the power requirement per ton.

This formula has been used successfully by the writer for years for estimating the loading or feed rate of mills having a length to diameter ratio of 2 to 1 or less, and operating in closed circuit with an efficient classification system. For longer mills a correcting factor to increase the loading should be applied.

On the other hand, it is known that shorter mills exhibit better flow characteristics and often contribute to less overgrinding. Therefore, from a process point of view, it is not acceptable to select a mill simply on a power basis leaving the mill dimensions and specifications to the design engineers.

CLASSIFICATION

The role of classification is to remove from the grinding circuit the particles fine enough or liberated enough to be processed.

A reduced classification efficiency results in an increased production of slimes which increases the consumption of flotation reagents, reduces the mineral selectivity and the recovery, creates materials handling problems in thickeners and filters and causes waste of time and equipment.

The efficiency of the classification system affects also the viscosity of the circuit and the flow through the grinding mills.

Some years ago, the standard classification system consisted of two stages of screw or rake classifiers equipped with sand spray washing.

In too many plants today this efficient system has been replaced by a single stage cyclone fed directly by a fixed speed centrifugal pump. This is a typical example of capital and space savings at the expense of process efficiency.

A cyclone is essentially a low efficiency classifier, and when fed directly by a constant flow pump it is an ideal setup for classifying finer on harder ores and coarser on softer ores. Even when equipped with sophisticated instrumentation it is a low efficiency system; after all a Chevette all equipped never makes a Cadillac.

Do you know anybody who has succeeded to control both ends of a cyclone at any one time? In fact very few people can control even one end at a time! Mind you there is nothing wrong with cyclones as part of a classification system; but a cyclone by itself does not make a system. At least if the cyclone were fed by gravity, the installation would permit some control; the dilution at the pump box could be set constant and the pump allowed to surge during reduced feed tonnage. The level in the cyclone feed tank would then adjust itself with the variation of feed tonnage.

Tests conducted in various plants have also shown that the overflow of gravity fed cyclones exhibits less cyclic size structure variations than the pump fed cyclones. In this latter case the variations are caused sometimes by pump box sand build-up.

If the classification system is built around cyclones one or both cyclone products should be cleaned, to reach an acceptable efficiency. For example the cyclone could be adjusted to produce a reasonably clean underflow and the dirty overflow would be screened or recycloned. With normal cyclone adjustment the underflow could also be further washed and decanted in a screw or rake type classifier, or in a counter current washing cone.

It must also be remembered that any type of gravity classification will promote a preferential grinding of the heavy minerals. If the associated middlings have a lighter density as with most iron ores, gravity classification may be used only for primary or rougher classification but with fine screens for final classification.

Admittedly, an elaborated and refined classification system will cost slightly more to operate but normally will generate higher profit.

Some flexibility should be built into the design of classification systems and generous sizing of key process units and materials handling equipment is important.

PROCESS

The process design should be as simple as possible and a wise mineral processor will apply the following principle: "When you do something, do it right the first time", and eliminate those recirculations that waste time and equipment and often are just an excuse to justify oneself. To meet this objective each operation must be properly equipped, the optimum operating conditions must be defined, the type of equipment selected best suited for the job and the layout should allow for proper maintenance. It is not in the scope of this paper to analyze in details how this principle can be applied, but as an example let's have a closer look at a flotation process.

Firstly, admit the hypothesis that there is no such a thing as a slow floating mineral, but rather poor chemistry, or inefficient equipment, or combinations of both; then recognize that the duty of a flotation cell is not to agitate the pulp but to introduce air bubbles in the pulp, and if this can be made with a minimum of agitation so much the better particularly when you are floating weak attachment minerals, like silica or spodumene, or when you are treating ore containing soft minerals, like calcite, which will generate large amount of slimes upon scrubbing or mechanical agitation.

The Davcra and Maxwell cells are valid developments in that respect and there is place in some flotation circuits for cells like South Western Engineering or deep air cells which have been discarded and forgotten too easily. Without the use of these types of air cells, the flotation of spodumene, at 35 mesh grind, at Quebec Lithium would have never been succeeded, neither the direct flotation of pyrochlore from carbonatite ore.

It is difficult to understand that none of these air cell types have been investigated in plants floating silica, while it is well known that the turbulence created by mechanical cells has to be reduced to the minimum through speed reduction to succeed in the flotation of 100 mesh quartz particles.

On the other hand, the various types of mechanical flotation cells on the market may exhibit a similar overall efficiency on a given ore, but often one type may well be more efficient in a finer size range, while another type will be more efficient in a coarser size range. In this case, the optimum results will be obtained by a combination of both types; and if this is not done for standardization reasons, it is a typical case of sacrificing the metallurgical efficiency for maintenance purposes.

In summary, the point here is that if the equipment is properly selected, and the various treatment stages properly equipped, the recirculation will be minimized along with the amount of equipment and the space requirements.

One cannot refrain from pointing out the flotation circuit of Falconbridge Copper at Chapais as a good example of simplicity, compactness and efficiency.

A process treatment is like a chain; and "a chain is never stronger than the weakest link". If there is for example a thickener in the circuit, this piece of equipment will accumulate mainly fine material during a flow increase and deliver that accumulation during a reduced flow. On account of this the size structure of thickener underflow material will be disturbed during flow variations, and this could adversely affect a following process sensitive to size structure distribution, e.g., pelletizing. In such a case an agitated surge storage ahead of the thickener, to absorb those flow variations is required to protect the efficiency of the following process.

INSTRUMENTATION

The potential benefits and usefulness of instrumentation in new concentrator design is even greater than for existing plants, as the design can be arranged to better accommodate the sensors, the measuring and recording devices.

The plant operating metallurgist must optimize the profits, and one way of doing it is through "total management" which implies: complete information, accurate provision and just decision.

Any change in the operating conditions must be detected and corrected; process modifications must be performed to compensate for predicted changes in operating conditions. All operating data must be registered and analyzed to determine the possible corrections.

Because a plant is usually treating the same ore, coming from the same deposit, one could question the magnitude of these changes, but W.J.Barkley, Superintendent of Operations at Kennecott Copper Corp. stated a few years back:

"The ore of Kennecott varies greatly in hardness, grindability, mineralization and moisture content. These variations affect the material handling, the crushability, and the grindability to the point that critical variations of capacity arise. The grade and tonnage of concentrates can vary up to 300%. These variations seriously affect the efficiency of the concentrator and of the smelter".

This statement illustrates the importance of a good control of the operations, and points out the metallurgical benefits and the operating cost reduction which can derive from such a control.

Detection, notation and correction of any change in operating conditions or results, a prime role of the operator, can be better performed with the support of instrumentation.

The operator, busy adjusting the circuit will postpone until later (and usually until the end of the shift) the notation of operating data with the result that even with the best intentioned, the report will be incomplete, making complete data analyses impossible.

There could be also many possible circuit corrections and a precipitated action may result in a loss of control, while the instrumentation will precede from a preset mathematical program.

To establish at the start a good program of instrumentation and automation, the process must be well known; the variables to be controlled must be identified along with the optimum level of these variables.

The laboratory and pilot plant tests would indicate and identify the variables but the true optimum operating conditions must be determined by tests in the plant itself. Moreover, the optimum operating conditions are not necessarily the ones that give the best metallurgical results but the best profits.

Automation does not replace operators but helps them. It is not yet generally applied in small plants but well accepted in large capacity plants which can more easily justify a sophisticated automation, because a slight increase in recovery represents in their case, a large saving compared to the capital and operating costs of controls.

The use of automated process controls is bound to increase. Many sociological, economical and technical factors, namely the increasing number of technicians on the market, the labour costs and the need of better control to treat more complex and lower grade ores induces the trend in that direction.

SUMMARY

Future concentrator plant design will have to pay more attention to energy consumption and environmental considerations. These will tend to increase the capital cost requirements.

To compensate, you will probably see more unhoused primary crusher installations like at Similkameen, more integrated crushing-grinding plants like at Niobec, more simple and compact flotation plants like at Falconbridge Copper, more large equipment installations, and better instrumentation.

For mineral processors, however, the key to improved future plant design and operation is better metallurgy which can only be the result of improved research and development. You do not have to be employed in a laboratory to do research and development; it can be done in plant operation, by looking at the way of doing things with a critical eye, trying every day to improve the results of a particular operation, and performing the necessary plant modifications to apply your innovative ideas.

Recognize that your metallurgical process can always be improved and you will be contributing to better, simpler and less costly plant design and operation.

REFERENCES

1.	THORNTON, E.	"The Effect of Grinding Media on Flotation Selectivity".
2.	FAHLSTROM et al	"Autogenous Grinding at Boliden".
3.	BRUCE, R.W.	"The Effect of Grinding Media on the Selective Flotation of a Copper-Zinc Ore".
4.	PETRUK, Wand Hughson M.R.	"Image Analysis Evaluation of the Effect of Grinding Media on Selective Flotation of a Zn-Cu Ore".
5.	SPENCER, A.G. and TAYLOR, A.D.	"The Influence of Large Unit Equipment on Mine Plant Design".
6.	BOLLES, J.L.	"The Morenci Concentrator Process Control System".
7.	COLEMAN, Richard L.	"Metallurgical Testing Procedure".
8.	KRIST, E.E. and SWEETSER, W.H.	"Anamax Mining Co. Orebody Sampling and Analytical Methods.
9.	LEWIS, COBURN and BHAPPU	"Comminution: A Guide to Size Reduction System Design".
10.	MACPHERSON, A.R. and TURNER, R.R.	"Autogenous Grinding from Testwork to Purchase".
11.	RIETHMANN, R.E. and BUNNELL, B.M.	"Application and Selection of Spiral Classifier".
12.	MULAR, A.L. and JULL, N.A.	"The Selection of Cyclone Classifiers, Pumps and Pump Boxes".
13.	ARBITER, N., HOBSBAWN, P.B. MAHONEY, J.F. and HARRIS, C.C.	"Conceptual Design of Flotation Circuits".
14.	SHOEMAKER, R.S. and GOULD, W.D.	"Modern Mill Design".
15.	RIETHMEIER, W. and LYNCH, A.J.	"The Future of Plant Design in Mineral Processing".

Design and Operation of a Future Mill

D. E. Pickett, P.Eng.

The Canadian Mineral Processors is particularly well prepared for a discussion of design of a future mill since in recent programs you have probably discussed all the present technology that will be the basis of future mill design and operation. Last year's session on energy utilization is a good example. Current literature has also provided ample coverage of the subject, the most recent and relevant being the AIME publication Mineral Processing Plant Design (1).

For future design we must also look to probable or possible results of development and experimental programs now in progress. Finally, as future design will depend on future knowledge, we can speculate about future research results which, hopefully, will add new possibilities to be developed.

Perhaps the best place to start our discussions would be a list of some of the future conditions that will affect mill design in the future.

(a) Costs will be even higher relative to metal prices.

(b) Metal consumption will continue to increase.

(c) Ore will be lower grade and less accessible; e.i. mill feed will be more expensive per unit of contained metal.

(d) Energy will have to be conserved, regardless of cost.

(e) Environmental requirements will be more important.

(f) Political pressures will be more demanding, not only from government but also from labour and business interests.

To meet the challenges posed by these conditions mill design will have to:

(1) Keep costs as low as possible, and

(2) Improve metallurgy.

Costs

The principal factors that influence costs in milling and mill design are shown in Figure 1.

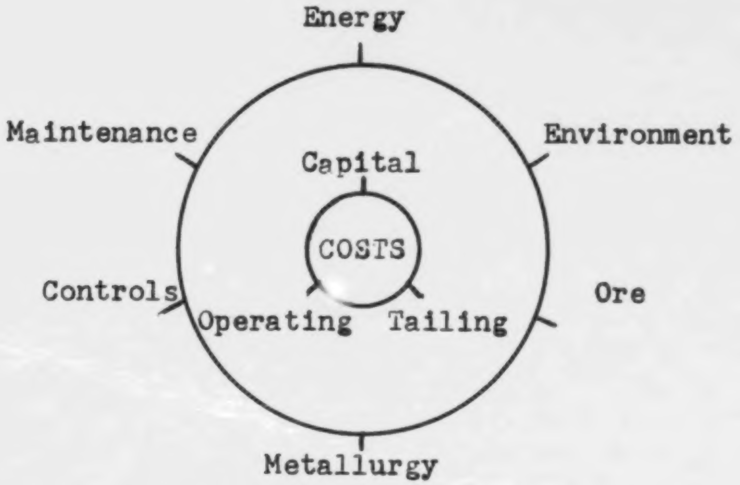


Figure 1 - Cost Factors

In 'Capital' we include not only costs of construction and equipment but also pre-production expense -- design costs, ore research and pilot plant expenses, etc. Under 'Ore' we include the mineral characteristics of the ore and also mining schedules, mine crushing, transportation, etc. **'Tailing'** is included as a cost in the same sense that L. E. Djingheuzian referred to "cost plus tails" as the real cost of treating an ore.

In a paper Modern Mill Design (2), Shoemaker and Gould state, "It is not well known in the area of plant design when, where, how and what to look for in order to end up with a minimum capital cost as well as a minimum operating cost." They list the most important considerations which include the following:

1. Use of largest possible equipment.
2. Minimum storage.
3. Provision for plant expansion.
4. Instrumentation.
5. Provision for maintenance of equipment.
6. Good working environment.

It follows that a discussion of each of these factors would be profitable, keeping in mind future conditions and future technology.

1. Use of Larger Equipment

The advantage of large equipment is said to be that there is a reduction in the number of parallel process lines (possibly to one line) bringing lower capital costs and lower costs of energy, labour, instrumentation and maintenance. However, possible problems are suggested in the area of equipment availability or productivity. Good design, especially design for easy maintenance, should compensate for these problems. Building design must provide for easy access and handling facilities for the larger and heavier equipment. To increase availability it should be possible to remove components or even whole assemblies for shop repair while reconditioned units are installed and in use. Material handling equipment, associated with the larger equipment, will have to be designed with care since breakdowns here will drastically affect availability.

2. Minimum Storage

Storage directly affects equipment availability, ore handling and equipment costs, capital costs, and metallurgy. Shoemaker and Gould state," Money spent for excess storage could be better put to use on purchasing more reliable processing equipment or in plant design which will permit operating availability."

3. Provision for Plant Expansion

In order to provide for plant expansion, future mining plans should be considered carefully before mill design, in so far as financial and political conditions can be forecast. Then the plant layout should provide for the probable expansion so that it can be completed with minimum capital and additional operating cost. The same consideration and provision should apply to other probable expansions such as installation of a computer, if not in the original design, or for development of by-product facilities as was done at Bernic Lake. We can only provide for probable expansion, not for all unforseen possibilities.

4. Instrumentation

It becomes increasingly evident that instrumentation can improve the performance of any mill. In small mills, even moderate expenditure on suitable instrumentation will reduce operating costs and improve productivity to a point where capital savings are possible in design. It is important to provide access for additional instrumentation as it may become available or necessary, not forgetting to provide access for sampling. Instrumentation should be chosen carefully and a good understanding of both the metallurgical characteristics of the ore and the operating characteristics of the equipment is essential. This requires special design capability and the cooperation of metallurgists, operators and maintenance experts.

For large mills computerization of this instrumentation will have important benefits. The success of the Brenda computer control system is recorded in the CMP paper by Bradburn and co-authors (3) . The Brenda experience demonstrated that:

(1) Grinding can increase production. (by 10% at Brenda)
(2) Power savings are possible. ($24,000 pa)
(3) Flotation recovery can be increased. (0.5% Cu and 1.0% Mo)
(4) A stand-by computer can be used for research and management.

The experiences at Mattagami Lake Mines (4), Kidd Creek, Falconbridge, INCO, and others have been reported in CMP proceedings, proving that well-designed computer control is essential in future design of large milling plants.

5. Provision for Maintenence of Equipment

Maintenence can be the major component of operating costs and is certainly the major problem for the mill superintendent. Rushton (5) claims that, "Most new plants never reach design production during the first year... or even five years .... Adequate consideration of maintenance functions usually fails during engineering and construction..." He adds that " technology is available to build a good plant anywhere... but the inclusion of experienced maintenance people in the design process is necessary." At the very least we should put more good design and less 'chiselling' into the design of material handling equipment.

Energy

In the sessions on energy, Joe (6) pointed out the importance of considering energy cost as a fraction of the final cost of the metal product. His summary also revealed that the major energy consuming operation in the mill is comminution which uses as much as 50% of total mill energy. Energy for drying is also high (30%) so drying should be avoided in design if at all possible.

Doyle (7) showed that there are very important energy saving possibilities in load management, and instrumentation to enable this should be included in plant design. Buildings will have to be better insulated with coordination of heating and ventilation requirements. This also applies to lighting requirments; it may be possible to reduce general mill lighting levels while providing access to plugs for area spot lighting and temporary lighting for maintenance.

Crushing and Grinding

The comminution area is the most costly from all aspects so has been the subject of a great deal of design prediction and published technology (8, 9, 10, 11, 12). Riethmeier and Lynch (10) state that " in the future, the total comminution system will be evaluated including crushing,grinding, screening, and classification, all considered in conjunction with the final concentration process." Probably the greatest benefit will be obtained by a combination of better metallurgical calculations as proposed by Steane (11) and better controls as described in the papers on the Brenda and Mattagami operations. If, as shown at Brenda, computer control can prevent undergrinding fluctuations and, as shown by Steane, a 19% saving can be obtained by moving to a coarser set point, overall savings of almost twice the '19%' should be safely achieved. This is of the same order as the 30% saving claimed possible by Flavell (8).

Crushing equipment will be larger to eliminate material handling where the greatest loss of availability occurs. Better equipment, better maintenence, open circuits, better sizing, better design of storage facilities, and generally better design will raise the expected availability of crushing plants from the 70 - 75% suggested by Steane to something nearer 90%. Better conveyor availability is possible as shown in the Mining Engineering paper on conveyor maintenance (13); e.g. if the most serious problems arise from inadequate supporting structures, the design remedy is obvious.

It appears that both 'autogenous' and'conventional grinding will be used in the future. With the substitution of large ball mills for rod mills in the primary stage, the restrictions on mill size will be removed in conventional grinding. However, in many cases the best combination may be autogenous or semi-autogenous primary grinding thus eliminating some crushing with ball mills for final separation in the secondary grinding stage. It may be possible to insert an intermediate primary separation or cobbing stage after the autogenous grinding to reduce the tonnage to be finely ground. All these predictions are feasible only if the ore characteristics and metallurgy will allow. Such characteristics as gangue liberation, stickiness, mineral segregation or dispersion will be important factors.

Mineral Separation

Separation processes will include more gravity, magnetic, electric and other separation methods although flotation will continue to be the most important in base metal mills. The obvious areas to investigate are the use of larger equipment, computer control and concentration at a coarser size to reject gangue as soon as possible. Arbiter and co-authors (14) in their predictions on flotation plant design state, "It is in the flotation circuit that the greatest potential for maximizing gains probably rests."

Finishing Circuits

There will probably not be many changes in the equipment in the circuits following the primary separation stage. Regrinding, desliming and all other possible unit processes are now being used. Design errors such as in the concentrate bins at Fox (15) should be discouraged.

Environmental Considerations

Better plant environmental conditions including safety, noise abatement and generally good housekeeping will pay off in better operation, better control and better labour relations.

The external environmental considerations are beginning to include not only ways of safe disposal of all plant effluents but also public and political relations. These are certain to cost more in the future. Riethmeier and Lynch (10) predict that a major effect of environmental regulation will be to "stretch project development schedules ... maybe ten to fifteen years" and increase costs accordingly.

Metallurgy

Although cost considerations as outlined above will be foremost in future design plans, metallurgy will also have to be improved if future design is to meet its objectives. It has probably occurred to many of you that most of these cost controlling proposals -- larger equipment, better instrumentation, improved design for maintenence, good plant environment and model effluent disposal practice. However, this may be because cost pressures have been greatest in these areas and thus 75% of research and development in the last 20 years has been directed toward cost controlling objectives with less than 25% toward new and better metallurgy.

In mill metallurgy we have made only a few major developments. Also, with few exceptions, these have been imported, e.g. pelletizing and flotation of iron ore from USA, Bartles-Mozley tables from UK, molybdenite flotation from USA, high intensity magnetic separation from both USA and UK. Several years ago the late Professor Wolfe of U of T wrote a paper "Has Your Mill Had a Check-up Lately?" We should ask now, "Has Your Metallurgy Had a Check-up Lately?" Costs are important but metallurgy is where the 'pay-off' is. It is the heart of mill operation.

Evans (16) has proposed a cost-benefit procedure which, if applied to metallurgy, could point out opportunities for research as well as design. Steane's calculations show that benefits can be obtained from a cost standpoint but we must look at possibilities for product improvement and reduction of tailing losses if the overall cost of metal production is to be reduced. A good example of the effect of a metallurgical change on metal economics is to be found in Kakela's paper (18) on the economic effect of iron ore agglomeration on the cost of iron.

Possible targets for research would include procedures and equipment for 'coarse ore' concentration. We need more research on magnetic separation, electric sorting and gravity processes, not research to find out how things we already have work but research into new methods and new equipment. George Riley's dry jig is a good example.

Currently, there are research projects in solvent extraction that may yield 'fall-out' ideas for more selective flotation reagents. Hydrometallurgical techniques have been recommended by Gow (15) and in this area the objective whould be to develop new operating procedures and new equipment, preferably larger, that will help in the up-grading of mill products and overall extraction.

In the long run, if the metallurgy is designed to reduce mill complexity, the plant design will be better and less costly. We should look closely at circulating loads and particularly at separate middling retreatment circuits. If computer control is to be used to optimize separation circuits it will be necessary to separate the primary (rougher) stage from the secondary (up-grading) stages. Economically, the rougher stage (and comminution) is related to the mining operation on a cost-per-ton-of-ore basis. On the other hand, the cleaning and concentrate up-grading processes, as well as gold and uranium refining circuits, can be more meaningfully related to the cost-per-ton-of-product. This will present opportunities for cost-benefit analysis, better control, effective research, and ultimately better design.

Finally,the importance of preliminary research, bench testing and pilot plant development prior to design cannot be overstressed if the metallurgy is to lead to good design. Design, without ensuring that the variability of the ore can be accomodated by the metallurgy to produce an acceptable result, is asking for trouble -- high costs, high tailing losses and, ultimately, expensive re-development, re-design and reconstruction. There are numerous examples, some resulting in failure of the mining venture, some only now working their way to a successful conclusion. Such design will be even less acceptable in future conditions.

The most pressing need for the future is mineral processing research and the most benefit will come from research, as in design, where there is full participation by the operating and design metallurgists in identifying research needs and in putting the results to practical use.

References

(1) Mular, A. L. and Bhappu, R. B. - Mineral Plant Design, AIME Symposium on Mill Design, 1978.

(2) Shoemaker, Robert S. and Gould, Wayne D., Modern mill design, Ref.(1), p. 826

(3) Bradburn, R., Gratch, E., Larsen, C., and Flintoff, B., Application of a process control computer at Brenda Mines Limited, CMP Proceedings 1978, p. 96.

(4) Konigsmann, K. V., Hendricks, D. W. and Daoust, C., Computer control at Mattagami Lake Mines, CMP Proceedings 1976, p. 173.

(5) Rushton, John W., Maintenance considerations in the design of a new plant, Mining Engineering, AIME, Mar. 1978, p. 253 and Apr. 1978, p. 364.

(6) Joe, E. G. , Energy consumption in Canadian mills, CMP Proceedings 1978, p. 15.

(7) Doyle, D. M., Energy conservation at Falconbridge Nickel Mines Limited, CMP Proceedings 1978, p. 56.

(8) Flavell, Malcolm D., Scientific methods to design crushing and screening plants, Mining Engineering, AIME, July 1977, p. 65.

(9) Lewis, Milton F., Coburn, James L. and Bhappu, R. B., Comminution - a guide to size-reduction system design, Mining Engineering, AIME, Sept 1976, p. 29 and Nov 1976, p.30.

(10) Riethmeier, Walter and Lynch, A. J., The future of plant design in mineral processing and extraction metallurgy, Ref. (1) p. 841.

(11) Steane, H. A., Some crushing and grinding considerations in plant design, CMP Proceedings 1976, p.15.

(12) Jones, R. E., The sizing and operation of crushing and screening plants, CMP Proceedings 1974, p. 141.

(13) Anon., How to maintain your mine belt conveyors, Mining Engineering, AIME, Mar 1978, p. 257.

(14) Arbiter, N. Hobsbawn, P. B. and Harris, C. C., Conceptual design of flotation circuits, Ref. (1), p. 447.

(15) Sharp, M. V., Chlopas, J. and Barratt, D. J., Improving the material flow in the Fox Mine concentrate bins, CMP Proceedings 1978, p. 317.

(16) Evans, J. B., Operating decisions based on cash flow analysis, Ref. (1), p.181.

(17) Gow, W. A., The role of hydrometallurgy in mineral processing, CMP Proceedings 1974, p. 284.

(18) Kakela, P. J., iron ore: Energy , labor and capital changes with technology, AAAS Science, Vol. 202, no. 4373, Dec 15, 1978, p. 1151.

Discussion of paper by R. Faucher and D. Pickett

E. Joe: What mechanism does an operator have to feed his information to the manufacturer for future mill design? At Outokumpu Oy the State operates mines and refineries and also builds equipment so there is an immediate interface. However, this is not the case in Canada.

D. Pickett: This is a big problem and it has been recognized. Mr. Stemerowicz has told me that in his experience, the consultants never came back to see how a mill worked and therefore, the mistakes from the previous mill were repeated in subsequent mills. The consultants will have to go back and not see what is right but see what is wrong.

M.C.Campbell: I had the opportunity last week to talk to some of the manufacturers of mine and mill equipment in Canada and they feel excluded from the CMP but would like to be in partnership. I think that the CMP should consider inviting the technical representatives of appropriate manufacturers of mine and mill equipment to the meetings.

THE AFTON CONCENTRATOR

JOHN LOVERING - MILL SUPERINTENDENT

MICHAEL ALLAN - MILL METALLURGIST

THE 11TH ANNUAL MEETING

OF

THE CANADIAN MINERAL PROCESSORS

OTTAWA, ONTARIO

JANUARY 23, 1979

THE AFTON CONCENTRATOR

INTRODUCTION

The Afton Mines property is located 180 air miles east-northeast of Vancouver and nine miles west of Kamloops, a city of 60,000 people in south central British Columbia. The deposit is located adjacent to the Trans-Canada Highway at an elevation of 2,231 feet above sea level.

Interest in the property dates back to 1898, when a 330 foot shaft was sunk on the Pothook claim immediately south of the existing operations. Since that time many companies and individuals have investigated the possibilities of the area but it was not until 1972 that Afton Mines Ltd. located a significant zone of copper mineralization. The feasibility study was completed in the period 1974-1975, by Teck Corporation Ltd., who control Afton through 50% direct ownership and 65% ownership of Iso Mines, who own a further 23%. Construction began in March, 1976. Ore production started in August, 1977, and the first ore was fed to the mill on December 8. This was the third mine to be brought into operation by Teck since 1975, the others being Newfoundland Zinc and Niobec.

The orebody is a porphyry copper deposit which has undergone supergene alteration. It is 1,700 feet long in a west-northwest by east-southeast direction, 300 feet wide and dips at 55° to the south. The mineralization in the supergene zone, which accounts for two-thirds of the known deposit, is rather unusual in that the copper is present primarily as native copper and chalcocite. The grade is one percent with the distribution: 70% native, 25% chalcocite, and 5% chalcopyrite with bornite, covellite, and cuprite.

Following bench scale and pilot plant testwork in 1972 and 1973, the Afton mill was designed to treat 7,000 tons per day of one percent copper ore. Semi-autogenous grinding was selected on economic grounds and because of the expected difficulties in operating a crushing and screening plant with the relatively high clay content of the ore. The recovery was to be 87% in a circuit incorporating both flotation and gravity separation.

FLOW SHEET

Run-of-mine ore is dumped from 100 and 120 ton diesel electric trucks into a 42" x 65" Allis Chalmers gyratory crusher set at 9" closed side setting. The surge pocket, below the crusher, is emptied by a Hydrastroke feeder onto No. 1 conveyor which discharges onto the coarse ore stockpile. Total capacity of the pile is 200,000 tons, but the live capacity is only 25,000 tons, sufficient for roughly three days mill feed. Six Hydrastroke feeders on two conveyors withdraw the crushed ore from the bottom of the stockpile. These two conveyors in turn discharge onto the belt feeding the semi-autogenous mill. An electronic scale on this belt continuously weighs mill feed.

First stage grinding is accomplished in a 28' diameter by 12'long Hardinge Cascade mill, driven by a 5300 HP direct current, variable speed motor. The mill is in closed circuit with a 50" Krebs cyclone and currently the mill is charged with 4" diameter balls to 10% of the mill volume. The mill liners are chrome-moly alloy with single piece rail liners on the shell and wedge bar lifters and plates on the ends. The grates are 5/8" relieved to 1½" without ball ports.

The second stage of the grinding circuit consists of a 16'6" diameter x 29' long Hardinge overflow ball mill, driven by a 4600 HP synchronous motor through an air clutch, in closed circuit with a Krebs Cyclopac containing ten 20" cyclones. The secondary cyclone overflow, at 35% solids, is flotation feed. The grind is approximately 75% -200 mesh. The mill is rubber lined and charged with 2" balls.

Gravity concentration begins in the grinding circuit where a portion of the secondary cyclone underflow is fed to 24" x 36" Denver Duplex Mineral Jigs. The jig tailing goes to secondary mill discharge while the concentrate is dewatered in a spiral classifier and conveyed to the 9'6" diameter x 12' long Hardinge regrind mill. This mill is also rubber lined, powered by a 500 HP motor and driven through an air clutch. The discharge from the mill is cycloned in 10" Krebs cyclones and the underflow feeds four triple deck Deister shaking tables. Table tailing is recycled via 15" Krebs cyclones with the underflow returning to the regrind mill. The overflows from both sets of cyclones report to cleaner flotation. Table concentrate assaying 85-90% copper, is filtered on two Straight Line horizontal belt filters and conveyed to a storage bin inside the mill. Approximately 30% of the total copper is recovered in this concentrate. However, operation of the gravity circuit is only necessary when the ore contains relatively coarse native copper.

FLOW SHEET - cont'd

The flotation circuit consists of 16 - 600 cubic foot Denver rougher-scavenger cells followed by 8 - 300 cubic foot Denver cells for first stage cleaning and 24 - 100 cubic foot Denver cells for the remaining three cleaner stages. Flotation concentrate at 50% copper or better is pumped to the smelter for thickening, filtering, and drying. Typical daily production is about 90 short tons, accounting for approximately 60% of the copper in the feed. The tailing is pumped through 6,730 feet of 18" O.D. Sclair pipe around the northern edge of the pit to a drop box and then flows by gravity to the tailing impoundment area. The two dams are being constructed according to the specifications laid out by Klohn Leonoff Consultants Ltd. and, since they are designed for water retention, there is no necessity for distributing tailing along them.

PERSONNEL

The mill staff consists of a Superintendent; a General Foreman with four Shift Foremen; a Metallurgist with a Junior Metallurgist and two Technicians; and a Chief Chemist with ten Assayers.

The assay office is part of the mill department but in addition to mill assays does work for the mine, smelter, and environmental engineering departments. Assayers work shifts to provide control assays for the mill and smelter. Between 8,000 and 10,000 determinations are carried out each month and approximately 20% of them are performed for the environmental monitoring program.

The operating crews under the Shift Foreman have three Operators and two Helpers. Since July they have been operating on a 12 hour shift, on a four and four basis. A typical eight day cycle consists of two 12 hour day shifts (8:00 AM - 8:00 PM), 24 hours off, two 12 hour night shifts (8:00 PM - 8:00 AM), and then four days off. We have now completed the initial six month trial period and the schedule has proven to be extremely popular with everyone involved.

START-UP

Mill operation commenced on 8 December, 1977, after several unexpected delays, the most serious of which was caused by a faulty transformer on the primary mill drive. A temporary replacement was found and the mill started up and ran for over two months on the replacement until the original transformer was repaired.

The mill start-up was hampered by minor problems in the fresh water pumping system, and difficulties in withdrawing material from the stockpile which had been started in August and by December had compacted to the point where it would not flow. Following several unsuccessful attempts using compressed air, water and explosives the pile was bulldozed flat and rebuilt with freshly crushed ore.

The ball charge in the primary mill was initially 2% of mill volume and was gradually increased to 8% to obtain the desired throughput. The initial charge consisted of 50% 3 inch and 50% 4 inch forged steel balls. A graded charge of 280 tons of balls from 1¼ inch to 2 inch was added to the ball mill, with additions of 2 inch balls as required. Design throughput was achieved within two weeks of ore being fed to the mill.

OPERATION

The concentrator is subject to extreme variations in feed with grades varying between 0.28% and 2.8% copper on a shift basis. This is a consequence of the lack of flexibility available to the pit operation due to equipment size and relatively few ore faces available at this time. The sizing of the pit equipment was dictated by the high stripping ratio, currently 7:1, and the need to provide large quantities of material for tailing dam construction.

Ore is crushed five days a week, one shift a day. There is no blending of ore except for that which occurs in the stockpile. Testwork prior to start-up indicated a Bond work index varying from 13.0 to 18.0. During the period of operation to date we have not had the opportunity to check this but we have experienced variations in throughput from 220 - 500 tons per hour.

OPERATION - cont'd

The mill is a single line plant which allows the use of larger equipment with lower operating costs. However, this advantage is offset, to some extent, by a lack of flexibility and the difficulty of evaluating the results of in-plant testwork, with rapid changes in ore type and no reference circuit available.

In an attempt to compensate for variation in feed the primary mill has been provided with a variable speed D.C. drive and it has proven to be extremely useful, especially during start-up. Speed can be varied manually from 0 to 78% of critical. The mill is normally run to draw maximum horsepower but can be slowed down to accommodate extremely soft ores which would otherwise overload the flotation and/or gravity sections of the circuit. This permits the tonnage to be reduced without damaging the mill liners. The variable speed has also been utilized for maintenance purposes and has the additional advantage of providing very smooth mill starts. It has resulted in very little liner bolt leakage, should give longer gear and pinion life and reduced liner consumption.

Notwithstanding the success of the variable speed drive, the grinding circuit has been the major problem area. The circuit is such that the primary and secondary grinding circuits are well balanced when treating the softer ore from the west end of the pit, but with the harder, lower grade ore, found towards the east end of the orebody, it has been found that the primary mill becomes the bottleneck while the ball mill is under utilized. A considerable amount of testwork has been carried out in an effort to rectify this situation. The 50" cyclones have been difficult to operate smoothly because of the variability of the ore and materials handling problems associated with the very coarse products in the primary grinding circuit. The intention is to set up one cyclone to classify at a coarser size in order to reduce the circulating load in the primary circuit when processing harder ores and thus restore the balance between the two mills. Changes to date have included pump speed changes and changes in feed line size to prevent sanding; installation of new lines on primary cyclone underflow and changes to cyclone variables such as inlet area, apex size, and vortex size. The test program is continuing but so far has not succeeded in reducing the circulating load below 100% and it has not been possible to coarsen the overflow to the range required. In spite of the attraction of running cyclones, from the maintenance viewpoint and for ease of operation, a decision was made to install a screen, on a test basis, in order to obtain more positive control of the size fraction returned to the primary mill and thus improve mill throughput on the harder ore and provide a slightly coarser feed to the flotation circuit. The installation was completed last week but there has been no opportunity to evaluate the results of the first few days of operation.

OPERATION - cont'd

Another change that has been necessary in the primary grinding circuit is the removal of the jigs, that were originally installed to treat a portion of the primary cyclone underflow. It was found impractical to operate them due to the coarseness of the feed.

The primary mill was designed to operate with up to 10% ball charge plus 25% rock by volume. Currently it is operating with a 10% loading of 4 inch balls and a total charge volume of approximately 13-15%. The ball mill has an overflow discharge and operates with the maximum (35% by volume) charge of 2 inch balls. Ball consumption to date has been 0.9 lb./ton in the primary mill and 0.8 lb./ton in the secondary mill. Large quantities of steel built up in the grinding circuit during the first few months of operation causing problems with pumping, making operation of the jigs difficult and adversely affecting the grinding efficiency. The situation improved after the worn grates in the semi-autogenous mill were changed. However, it is expected that magnetic removal of the steel will be necessary as the grates wear again.

The original testwork indicated that mill recovery would be about 87%. This was achieved in the third complete month of operation and since then has averaged 89%. The distribution of copper between the gravity and flotation concentrates varies with the nature of the ore. The productivity of the gravity circuit depends mainly on the coarseness of the native copper mineralization. Graph 1 shows typical copper distribution to the two concentrates by size fraction with flotation being generally more successful below 65 mesh. It has been found unnecessary to operate the gravity circuit when the jig concentrate falls below 5%, as the tonnage of table concentrate produced is minimal and there appears to be no improvement in recovery. There have been periods of as much as three weeks when no metallic concentrate was produced, while the record for a single day is 181.9 tons at 88.7% copper, which was 61% of the copper in the feed.

Flotation feed sizing is approximately 10% +100 mesh and 75% -200 mesh. Density is 35% solids. The original design called for cycloning the rougher concentrate and regrinding the coarse fraction along with the jig concentrate. This was found to limit tonnage by overloading the Deister tables. The rougher concentrate is now sent directly to the first stage cleaners and is upgraded in four stages to an average of 53% copper. Reagent consumption at Afton is high with current usage of collector averaging 0.25 lb./ton. Frother consumption was also 0.25 lb./ton until use of reclaim water lowered this to approximately 0.15 lb./ton. The original collector was Potassium Amyl Xanthate but plant testing has indicated Sodium Isopropyl Xanthate to be a suitable replacement for the majority of ore types and, as it is considerably less expensive, will be used whenever possible. Pre-production testwork suggested the use of a mixture of Cresylic Acid and M.I.B.C. as a frother and this was used until it was found that Pine

OPERATION - cont'd

Oil could be substituted for Cresylic Acid. The result was a superior froth, slightly lower cost, and less hazardous material to handle.

Gold and silver are present in the ore and with today's prices have had a considerable impact on the revenue to date. Mill heads to the end of June averaged 0.020 oz./ton gold and 0.148 oz./ton silver with recoveries of 79% and 63% respectively. The distribution of gold and silver between the two concentrates has been extremely variable but the average to date shows that 55% of the gold and 50% of the silver in the feed has reported to the flotation concentrate. Gold production from start-up to the end of September amounted to 32,973 ozs.

WATER SUPPLY AND TAILING DISPOSAL

The fresh water supply system is capable of providing 4,400 USGPM. Water is pumped a vertical distance of 1,400 feet from Kamloops Lake to the main head tank. Four 300 HP vertical turbine pumps in parallel are installed at the lake and four identical units are located in the booster station at a site selected to balance the pumping head at each station. The pipeline is 25,000 feet of 24 inch O.D. welded steel pipe. The booster pumps deliver water to the main head tank which in turn feeds the mill head tank.

The tailing pond is situated in the Hughes Lake valley, southwest of the pit and approximately 10,000 feet from the mill. Two major dams have been constructed with the tailing impounded between them. Both have been designed for an average height of 150 feet. Construction started in the summer of 1976 under the supervision of Klohn Leonoff Consultants. Each dam has a rock core with an impervious layer of glacial till on the upstream side. Between the rock and till are coarse and fine crushed rock, forming filter zones, which prevent the migration of fine materials. The upstream slope is 2:1 while the downstream will be benched with slopes of 1.5:1. A 20 foot high compacted earthfill dam below the west dam was constructed to handle seepage, but has not been required so far. There is no effluent from the impoundment area.

Original testwork indicated that settling of the tailing would be very difficult due to the bentonitic clay content. Consequently the

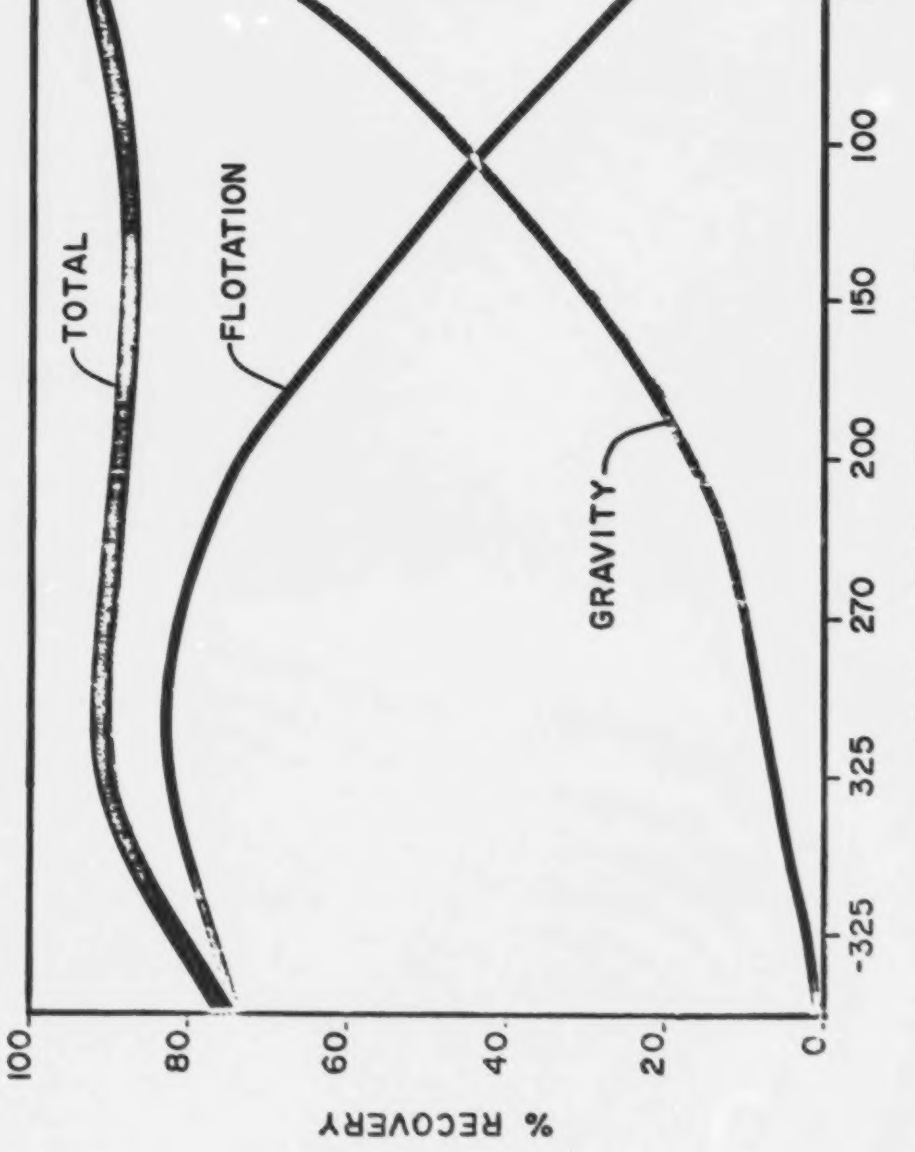
WATER SUPPLY AND TAILING DISPOSAL - cont'd

tailing dams were designed on the assumption that there would be no water reclamation and the total volume would have to be impounded. Fortunately this has not been the case.

Much work has been done in the laboratory to test various settling aids but so far it has not been necessary to treat the tailing in any way. Natural settling has produced a supernatant that contains about 300 ppm suspended solids which has proven to be acceptable for reuse in the mill process water. As a result a reclaim water system designed to recycle approximately 50% of the mill water requirements was installed during the summer of 1978 and is now in operation. Currently 35-40% of the water is being reclaimed. Further minor piping changes should permit utilization of the reclaim pumping capacity and allow closer to 50% reclaim. However, it is unlikely that the percentage of water reclaimed will ever exceed this because of the smelter's continuing requirements for cooling water which is directed to the mill circuit and thence to the tailing pond.

GRAPH I

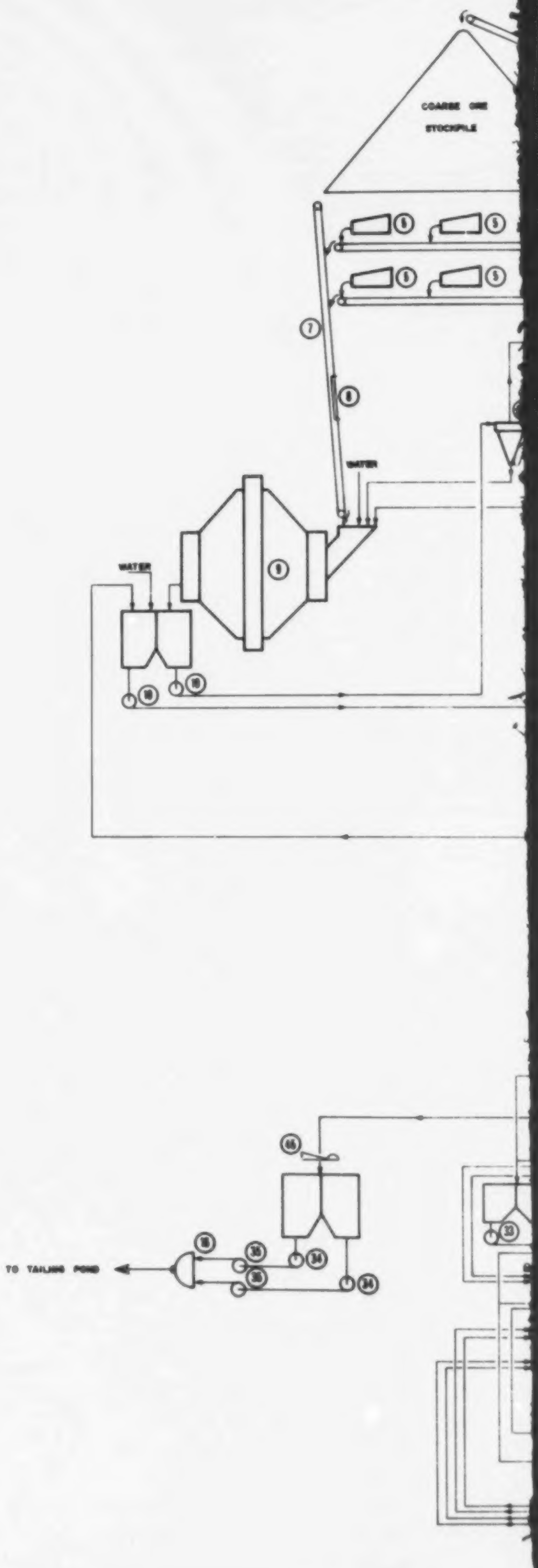
RECOVERY VS. SIZE FRACTION

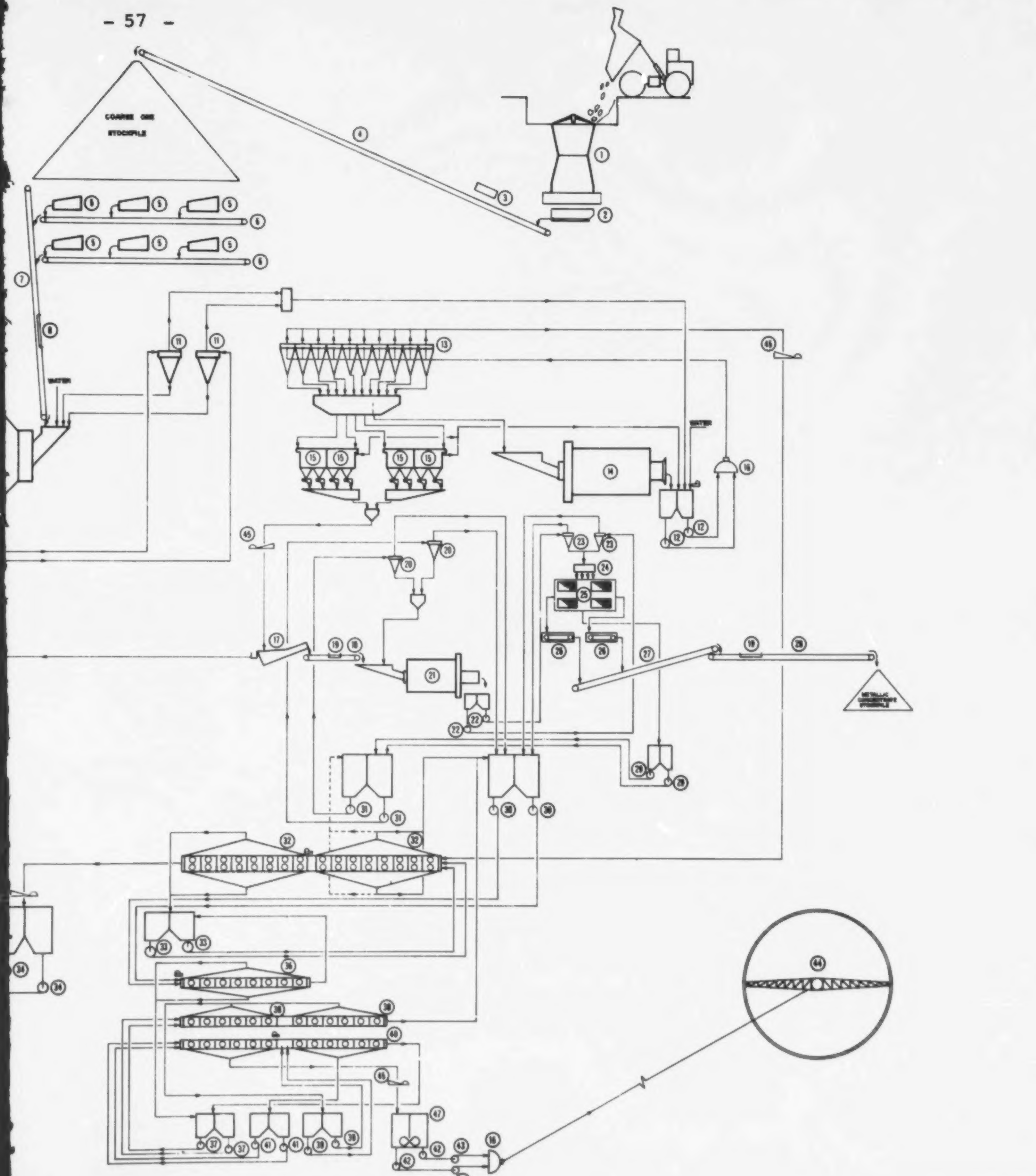


AFTON MINES LTD. (N.P.L.)
CONCENTRATOR FLOWSHEET

EQUIPMENT LIST

1 1- 42" X 65" ALLIS CHALMERS GYRATORY CRUSHER (350 HP at 900 RPM)
2 1- 72" X 16' NICO TR 1100 HYDRASTROKE FEEDER (5 SPEED)
3 1- AREA METAL DETECTOR
4 No 1 CONVEYOR (48" BELT WIDTH)
5 6- 48" X 10' NICO 550H HYDRASTROKE FEEDERS (VARIABLE SPEED DRIVE)
6 No 2 & No 3 CONVEYORS (42" BELT WIDTH)
7 No 4 CONVEYOR (42" BELT WIDTH)
8 MERRICK 480 DS 1 WEIGHTOMETER
9 28' X 12' KOPPERS PRIMARY MILL (5300 HP at 176 RPM) VARIABLE SPEED
10 2- 10 X 12 GEORGIA IRON WORKS PUMPS (PRIMARY MILL DISCHARGE) 250 HP
11 2- D50 KREBS CYCLONES (ADJUSTABLE APEXES) " "
12 2- 16 X 14 X 34 SRL C ALLIS CHALMERS 350 HP PUMPS
13 10- D20LB KREBS CYCLONES WITH PNEUMATIC INLET VALVES AND ADJUSTABLE APEXES
14 1- 16'6" X 25' KOPPERS SECONDARY BALL MILL (4800 HP at 180 RPM) SYNCHRONOUS MOTOR
15 4 24" X 36" DENVER MINERAL JIGS
16 3 TECH TAYLOR VALVES
17 1- 30" X 20'-3" TRIPLE PITCH WEMCO SPIRAL CLASSIFIER
18 No 5 CONVEYOR (18" BELT WIDTH) TWO SPEED DRIVE
19 2-MERRICK 440 DS-1 WEIGHTOMETERS
20 2- KREBS D15 CYCLONES
21 1- 9'-6" X 12 KOPPERS REGRIND MILL (500 HP at 1200 RPM)
22 2- 5 X 4 X 14 SRL C ALLIS CHALMERS 30 HP PUMPS
23 2- KREBS D10L CYCLONES (REGRIND MILL DISCHARGE)
24 1- DEISTER 12 WAY DISTRIBUTOR
25 4- CONCENCO 999 TRIPLE DECK CONCENTRATING TABLES
26 2 STRAIGHT LINE METALLIC CONCENTRATE FILTERS
27 No 7 CONVEYOR (18" BELT WIDTH)
28 No 8 CONVEYOR (18" BELT WIDTH)
29 2 5 X 4 X 14 SRL-C ALLIS CHALMERS 5 HP TABLE TAILING PUMPS
30 2 12 X 10 X 25 SRL-C ALLIS CHALMERS 30 HP REGRIND CYCLONE OVERFLOW PUMPS
31 2 10 X 8 X 21 SRL C ALLIS CHALMERS 60 HP REGRIND MILL CYCLONE FEED PUMPS
32 8+8 DENVER 600 H ROUGHER AND SCAVENGER FLOTATION CELLS (16- 60 HP at 900 RPM MOTORS)
33 2- 14 X 12 X 29 SRL-C ALLIS CHALMERS 50 HP SCAVENGER CONCENTRATE PUMPS
34 2- 14 X 12 X 29 SRL-C ALLIS CHALMERS 250 HP FLOTATION TAILING PUMPS
35 2- 14 X 12 X 29 SRL C ALLIS CHALMERS 250 HP VARIABLE SPEED FLOTATION TAILING PUMPS
36 8- DENVER 300-V CLEANER FLOTATION CELLS
37 2- 6 X 6 X 15 SRL ALLIS CHALMERS 15 HP FIRST CLEANER CONCENTRATE PUMPS
38 12- DENVER 30 UR CLEANER FLOTATION CELLS (6- 25 HP at 1200 RPM MOTORS)
39 2- 6 X 6 X 15 SRL ALLIS CHALMERS 15 HP SECOND CLEANER CONCENTRATE PUMPS
40 12- DENVER 30 UR CLEANER FLOTATION CELLS (6- 25 HP at 1200 RPM MOTORS)
41 2- 6 X 6 X 15 SRL ALLIS CHALMERS 15 HP THIRD CLEANER CONCENTRATE PUMPS
42 2- 5 X 4 X 14 SRL-C ALLIS CHALMERS 30 HP FINAL CONCENTRATE PUMPS
43 2- 5 X 4 X 14 SRL-C ALLIS CHALMERS VARIABLE SPEED 30 HP FINAL CONCENTATE PUMPS
44 1- 80' DIA FLOTATION CONCENTRATE THICKENER (AT SMELTER)
45 4- AUTOMATIC SAMPLERS
46 4- CLARKSON REAGENT FEEDERS
47 1 FINAL CONCENTRATE CONDITIONER TANK




COARSE ORE STOCKPILE
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METALLIC CONCENTRATE STOCKPILE
CD-2A

AFTON MILL SUMMARY DATA

	Design	Year to Date To Sept. 30/78	1978-79 1st Quarter Oct.-Dec./78
Tonnage (Cal. Day)	7,000	7,162	7,936
Recovery - Flotation	57	61	59
- Gravity	30	25	30
- Total	87	86	89
Flot. Conc. Production - TPD	100	82	85
- % Cu	50	53	58
Met. Conc. Production - TPD	30	21	29
- % Cu	90	84	86
% Operating Time	92	91	94

Heads (YTD) - 0.99% Cu
- 0.020 oz./ton Au
- 0.145 oz./ton Ag
} To September 30, 1978

Lbs. Cu Produced	34.8 Million	To September 30, 1978
Ozs. Gold Produced	32,973	
Ozs. Silver Produced	197,500	

Ball Consumption - Primary - 0.9 lb./ton YTD
Ball Mill - 0.8 lb./ton YTD

Reagent Consumption - Collector - Xanthate - 0.25 lb./ton
Frother - MIBC/Pine Oil - 0.15 lb./ton

Flotation Feed - Grind - 75% -200 Mesh

January 18, 1979

Discussion of Paper by J. Lovering and M. Allan

M. Robert:	I understand that the mill was designed to operate at 10,000 tons per day. Are You able to attain that ?
Reply:	Yes, during the last few months we have been averaging 8,000 tpd. This means that we are averaging 10,000 tpd on the soft ore. In fact, last Friday we made in excess of 12,000 tpd.
M. Robert:	At the same fineness of grind ?
Reply:	We are grinding a little finer than we should be. Probably on Friday it was about 65% minus 200 mesh. The feasibility study was based on that fineness.
W. Meekel:	What instrumentation do you have in flotation ?
Reply:	We have only level controllers. When we go to instrumentation we will concentrate on the grinding circuit before we go to the flotation section.
R. Tenbergen:	When you have a high percentage of native copper do you have any problem in getting this out of the mill or plugging of grates ?
Reply:	We have not had any plugging of grates but when the native copper is high we tend to get a build up of larger pieces.
J. Wallace:	Are you limited to using only 50% reclaimed water because of the water temperature ?
Reply:	The limiting factor is the smelter which uses water to remove mercury and there is a temperature limitation there.

K. Konigsmann:	What is the operating time on the primary mill ?
Reply:	94%. The feasibility study figure was 92%.

Paper No. 3

THE TEXASGULF CONCENTRATOR EXPANSION

BY

A.M. Hill - Project Engineer - Texasgulf Canada Ltd.

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THE TEXASGULF CONCENTRATOR EXPANSION

A.M. Hill, Project Engineer

This is an account of the expansion of Texasgulf's Timmins concentrator by the building of a fourth circuit.

The original capacity of the mine and concentrator was 3.6 million tons per year of ore. In 1974, Texasgulf decided to increase that capacity by about one third to 4.8 million tons per year. It was planned that the increased capacity would be available by mid-1978.

The expansion involved work at the minesite, increased railroad capacity, and a new circuit in the concentrator, and it is the concentrator expansion which is described here.

In 1974, the minesite drew its ore from the open pit. The pit was being phased out and was to be replaced by No. 1 mine. To allow for the expansion, a second mine was required. The railroad was capable of handling 18 car trains. Its capacity had to be increased to 27 cars per train and this, in turn, required two new locomotives. Each of these alone cost half a million dollars.

The concentrator building in 1974 contained the concentrating equipment and miscellaneous services (such as machine shop, warehouse, analytical lab). The grinding and flotation sections were divided into three circuits ('A', 'B', & 'C'). 'A' and 'B' circuits treated copper-zinc ore and 'C' circuit treated the much more difficult copper-zinc-lead ore. This 'C' circuit had been modified over the years to implement improvements in the metallurgy.

The task was to build a new circuit (called 'D' circuit) to treat the copper-zinc-lead ore, using the technology developed for it on the 'C' circuit, and afterwards, to convert the 'C' circuit to process straight copper-zinc ore. All these changes had to be made without interfering with production.

The overall cost of the expansion was $139 million, with $88 million allocated to the minesite expansion and $51 million allocated to the railroad expansion.

The concentrator had originally been designed with expansion in mind. The space for the new circuit was available within the walls of the concentrator but this space was being used to house various services which would have to be relocated. This relocation cost $14 million. The 'D' circuit itself cost $30 million and engineering cost $5 million.

Before construction could start, the various services had to be moved from inside the concentrator. First, the warehouse was relocated. The present spacious building cost $4 million and rationalizes all warehousing for the whole metsite. This released a zinc plant warehouse, which was turned into a new central changehouse for the zinc plant, offices for engineering, traffic and zinc plant, and new first aid facilities.

Next, the analytical laboratory was relocated. The new building, again, services the whole mine and metallurgical complex, bringing together concentrator and zinc plant laboratory facilities. The old laboratory at the zinc plant now houses the zinc plant development laboratory.

A brand new maintenance building was constructed at a cost of $3 million to house the centralized maintenance facilities for the whole site, and special services for all Texasgulf's Timmins operations. The old machine shop and railway equipment servicing facilities were relocated to this building.

In all, it cost $14 million for various relocations before a start could be made on the actual concentrator expansion.

Early in 1975, design work on 'D' circuit started. A team was assembled with a full time representative from the mill department and one from maintenance. The amount of design work needed was much more than Texasgulf's regular engineering department could handle alone, so Cambrian Engineering of Mississauga was engaged to supply the necessary manpower. Texasgulf appointed a project engineer to manage and co-ordinate all activities from design through to final handover of the completed facilities.

2600 new drawings were required. At peak, there were approximately 45 people working on the design. All drawings were scrutinized twice before final approval by the production and maintenance groups. This caused a real paper-flow problem but assured that all the improvements resulting from years of operating and maintenance experience were incorporated into the new design.

Even while the design was going on, as each section was completed, contracts were let for the work in the field. Each contractor was under the supervision of one of the engineering department's field supervisors. Work was done by local contractors with Texasgulf as general contractor.

The first step was to extend the crusher house and build a new pilot plant. Foundation work was done in the summer of 1975. Deep foundations were needed for two new crusher bases and the extension of the tunnel for crushed ore and recycle conveyors. The steelwork was in place and roofing and siding on before winter.

At the same time the road was excavated alongside the concentrator to bury two 30" diameter pipes to take raw water and blended water into the new pumphouse. The blended water pipe is made of fiberglass-reinforced polyester because of the potential for steel pipe to corrode. The FRP pipe is buried under the road, protected by a corrugated conduit. The service tunnel, which was installed from the new service buildings, but stopped short of the concentrator, was extended to link with the concentrator at this time.

The next steps in making room for the 'D' circuit were to relocate the pumphouse and to move the pilot plant and emergency generators. The new pumphouse represents more than a simple relocation. It incorporated all the improvements of previous years and also was upgraded to handle both 'D' circuit demands and those of the copper smelter and refinery. The shell of the pumphouse was built during the spring of 1976, and mechanical installation was completed by July. The startup of the new pumps involved making about a dozen 'live taps' into operating water pipelines. The new pumphouse went on-line in July without the need for any shutdown at all.

At the same time, the pilot plant equipment was being relocated to its new home. Again, this represented more than a mere relocation - the pilot plant was enlarged and its layout greatly improved. The concentrator department now has an up-to-date pilot plant of which it can be proud.

The emergency generators, too, were relocated to a new room at the other end of the concentrator. Some longstanding problems with the equipment were corrected at the time of this move.

And for the maintenance crews a small shop, and office were constructed.

By the end of 1976, all necessary relocations had been made in preparation for the actual 'D' circuit construction work inside the concentrator.

Concurrently, the concentrator handling and drying capacity was increased. The work in the load-out area was straightforward. It involved modifying various conveyors at the top of the load-out building to suit the revised product flows and to accommodate a new conveyor from a new dryer. Installation of the new dryer (value $1.5 million) was started in the summer of 1976. This work was sub-contracted out as a package to Nichols - the dryer supplier. The dryer came on-stream in May, 1977. The gallery for the conveyor, to take the dried concentrate from the bottom of the dryer to the top of the load-out building, was fabricated in Montreal and shipped in sections. It was erected in late fall and cladding was added during the winter. Then, the conveyor itself was installed.

Two additional vacuum filters were needed, which, together with supporting steel and concrete flooring, and many piping changes and extensions, were in place in time to feed the new dryer.

The load-in capacity had to be increased by 50%. This involved an extension of the facilities westward. While construction work was going on, the trains could not follow their normal procedure of pulling in, dumping ore, and continuing on in the same direction. A new (temporary) track pattern was needed which would allow loaded trains to bypass the load-in and split up and back in from the east in 9 car sections. The extra manoeuvring meant that the railroad crew had to haul on a 24-hour basis instead of 16 hours per day, which placed a strain on the traffic department and cut maintenance time to a minimum.

The business end of the load-in facilities is below ground level. For the expansion an excavation 50' deep at the west end had to be made. The law requires a maximum slope when excavating in soil of 45°, which, in turn, means that the top of the excavation was 175' x 130'. The edge of the excavation came close to the railroad track which is the lifeline for ore between the mine and the concentrator. Suddenly, water broke through in the north-west corner of the excavation and started to wash away the north side of the excavation wall - just where the tracks were.

Attempts to stop the water flow only forced the stream to another location and caused a further washout.

A round-the-clock program was formulated to support the north slope and continuously pump away the water. There were some anxious moments but the ore haul was never interrupted. There is now a permanent reminder of those few hectic days and nights in a continuously operating sump pump in the north-west corner of the load-in building.

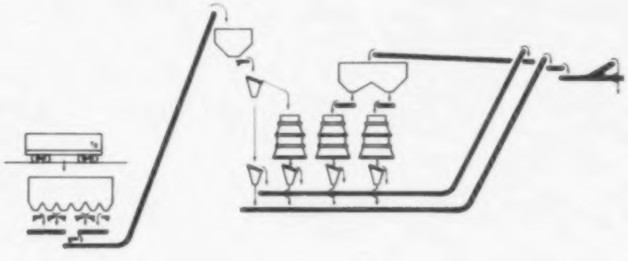
The load-in extension involved three new ore pockets, lined with heat-treated alloy steel, three new feeders, and the extension of the conveyor by 80 feet westward to take additional load and increase the capacity of the conveyor drive. The building extension, heating, ventilation, dust collection, instrumentation, electrics, lighting, and so on, went along with the major mechanical work.

The load-in feeds the crusher house via a conveyor. Its capacity was increased from 900 to 1200 tons per hour by increasing the motor size, incorporating a fluid coupling, and changing the belt. This, and about five other modifications in the crusher house had to be done 'on the fly'. Starting with full fine ore bins, it was estimated that the longest shutdown of the crusher house, before the bins were empty, was 30 hours. All shutdowns for major modifications had to be scheduled within this framework. They were always carefully planned and always started at midnight. Every shutdown was a success and the ore availability to the grinding circuit was never interrupted.

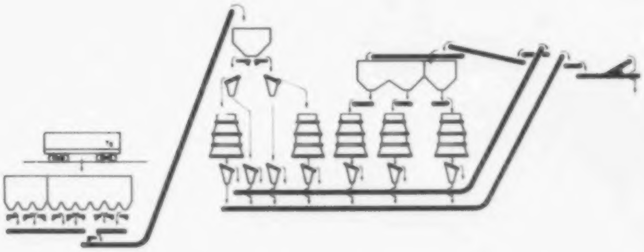
VIEW OF
CRUSHER
HOUSE
EXTENSION

Before the expansion, there were one secondary crusher and two tertiary crushers together with all their associated feeders, screens, conveyors, etc. To maintain uniformity of sizes and types of equipment, the new crusher house incorporated one new secondary and one new tertiary crusher with a total of four new screens and one new feeder. The conveyor modifications were more complex. They involved the extension and upgrading of two recycle conveyors: installation of a new transfer conveyor, a new shuttle conveyor, a new tertiary crusher feed conveyor with variable speed drive, and the extension and upgrading of the fine ore conveyor. All of this was accomplished with no production loss. It was here that the 'team' concept really paid off. The production people were never caught unaware by what the engineering department was doing - or vice versa.

SCHEMATIC
OF CRUSHING
CIRCUIT
BEFORE
EXPANSION



SCHEMATIC
OF CRUSHING
CIRCUIT
AFTER
EXPANSION



The particular problem with the shuttle conveyor is an interesting one. The old recycle conveyor discharged into a surge bin and thus fed either or both of the tertiary crushers. The height of the discharge point of the conveyor was enough to ensure that the crushers were fed evenly. But, with the addition of a third crusher, the elevation of the new conveyor discharge would have been so high that the conveyor slope would have been excessive. In any case, the headroom was not available and the bin load would have been too much for the structure. So, a new, slow moving, shuttle conveyor was selected to distribute the load between crushers in the extended surge bin. This system was plagued with problems from the start. Rocks bounced off dust seal supports under the drive wheels. The drive wheels, which were supposed to make the conveyor shuttle, slipped and it discharged all in one spot. The rubber dust seal along the open discharge port got tangled up in the drive axle; then, instead of the plow lifting the seal rubber, it got squeezed right under the chute. It took time and money to work out all of these problems, and a lot of brainstorming sessions. The solutions used were these. The drive axle was weighted to give traction; the rubber dust seal support bars and plow arrangement were removed and a piece of old conveyor belt was attached to the discharge chute which moves with it; and a deflector was installed to return bouncing rocks into the bin.

The crushed ore is stored in a fine ore bin, with a live capacity of 10,000 tons.

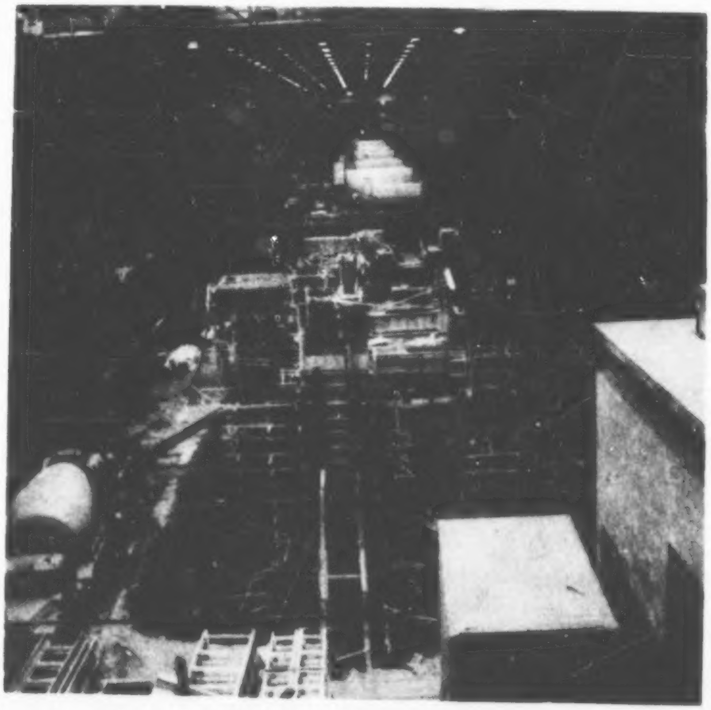
The foundations for this bin were put in the area generally vacated by the old pumproom, analytical laboratory, and emergency power room. Some massive steel went into the support structure and had to be hoisted into the building through the roof. This work was to be completed before winter set in in 1976, but, because work was a little behind schedule at that stage, it was decided to reschedule that phase of the work for the spring of 1977. And a good decision it was too, because the winter set in very early that year and the roof would have been off for at least a month of snow.

The ore is fed into the fine ore bins by a moving tripper conveyor. This conveyor was already 620 feet long and it had to be furnished with a drive of increased capacity, a new tail pulley, and a completely new belt (increasing the belt length by 320 feet) - all within the alloted 30 hours. So, now, fine ore can be delivered from the transfer tower to any of 'A', 'B', 'C', or 'D' bins.

The ore is withdrawn from the fine ore bins by seven slot feeders on to seven variable speed conveyors. These feed two collecting conveyors, which, in turn, drop on to the grinding rod mill feed conveyor. All these conveyors virtually duplicate the setup in the old circuits.

In the spring of 1977, work on foundations for the new grinding and flotation circuits was started. Some heavy foundations were needed for the grinding mills. When excavation started, it became apparent that bedrock was at higher elevations than expected. The mill foundations were redesigned in the field to take advantage of this. Foundations for four mills needed for 'D' circuit and for a possible fifth, future, regrind mill were installed.

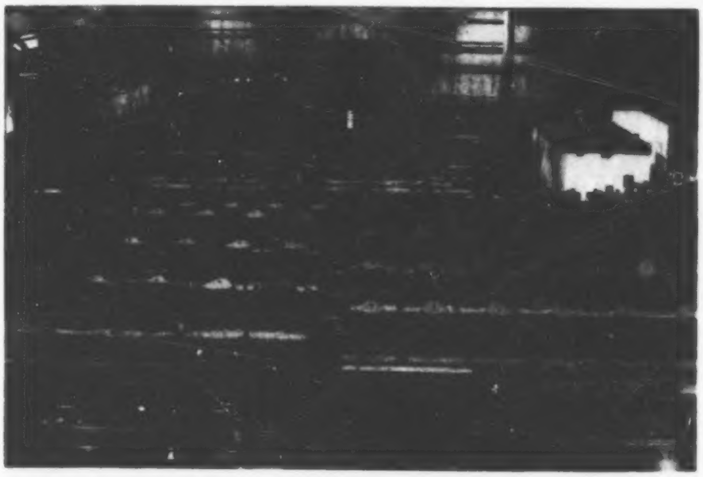
NEW GRINDING MILL FOUNDATIONS UNDER CONSTRUCTION



The mills themselves were brought in and installed using the existing maintenance crane. A representative from Allis Chalmers was on site throughout to direct the contractor on Texasgulf's behalf. Pumps, pumpboxes, cyclones, and all the associated piping formed part of the new grinding circuit. Here, as elsewhere, a multitude of small improvements which had been thought out over the years of operating and maintenance have been incorporated.

The flotation circuit was installed at the same time as grinding. In this area, too, Texasgulf's experience was brought to bear on the design. The layout makes this immediately obvious, especially below the operating floor level.

VIEW OF
NEW
FLOTATION
CIRCUIT



There are basically three flotation operations. The first flotation circuit produces copper, the second lead, and the third zinc. This zinc flotation needs a very fine grind, which accounts for the zinc regrind mill in the grinding area. Perhaps the most impressive numbers appear in the flotation area which included 13 km of pipe, 8,400 pipe fittings, 4,700 victaulic couplings, 750 valves, 160 launders, 2,300 litres of paint and 1/3 acre of Linatex.

To provide the greatest possible flexibility, the flotation circuit also incorporates the capability of treating the copper-zinc type ore. This was done quite neatly with the addition of one more bank of copper cleaner cells, and some extra distributors, pumps, and piping. So, the 'D' circuit is able to treat whatever the mine produces simply by switching a few pumps and valves.

The old 'C' circuit was modified so that it will handle only copper-zinc ore. This work was done to fit into a six-week shutdown of the circuit. After that, it again went into production, allowing 'A' and 'B' circuits to be taken off stream, in turn, for major overhauls.

Thus, by the end of 1978, three existing circuits had been completely overhauled, and, together with the additional fourth circuit, the concentrator was upgraded to its new capacity.

The hundreds of new motors and other gear needed power. And the task of doing this without ever interfering with production was, perhaps, the greatest challenge of all. The new power requirements with a connected load of 11,000 kW necessitated two new rooms - a high voltage and a low voltage switchgear room. The high voltage room contained two transformers, five high voltage starters, variable frequency inverter, and four motor control centres.

In the low voltage room were installed four 2000-KVA substations and eleven motor control centres controlling about 300 new motors. The whole of the electrical work involved 122 km of cable.

Texasgulf's concentrator is one of the most highly automated concentrators in the world and the newest circuit certainly advances even on that. In addition to the normal control panels to parallel the existing central control layout, a further feature was incorporated which allows the operator to call up on a CRT right at his desk a visual representation of the status of any one of his control loops. All the adjustments to circuit conditions, be they pH, density, or reagent additions, are handled automatically, and, what is more important, continually optimized to give the most economic setting. The computer is housed in a new room, vacated by the instrument shop. The instrument technicians have been relocated to a new shop, more spacious than their previous one.

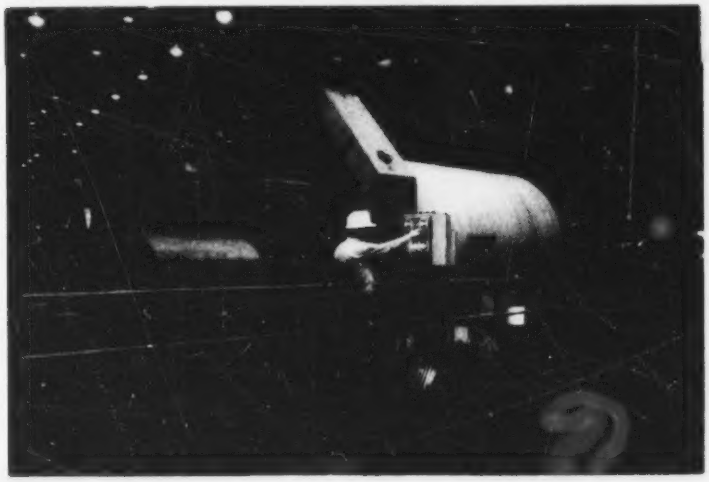
The story would not be complete without a mention of the extra load it placed on many of the other departments, which are not really staffed to handle a major construction project.

About 1,000 purchase requisitions were generated. Texasgulf's purchasing people had to send out enquiries, obtain bids, place orders and expedite. Warehousing needed a whole yard to store items on delivery. Accounting people had to keep track of payments on all these purchases. The maintenance department had to review the orders for spare parts and whole spare units. (Texasgulf invested $250,000 on capital spares alone.) They also helped with inspections of equipment installed by the contractors. Security and first aid people had added problems due to the influx of contractors' personnel - at peak, there were 230 men working on the project, representing well over 500,000 manhours. Safety was an ever present concern. Texasgulf is proud that the safety record of the contractors was at least as good as its own.

During the life of the project, there was hardly any group on the property that was not in some way affected by it, and that did not make a contribution.

Startup, six weeks ahead of schedule, went so smoothly that concentrate was being produced within six hours of putting ore into the fine ore bin.

STARTUP,
AHEAD OF
SCHEDULE,
PRODUCED
CONCENTRATE
WITHIN
SIX HOURS



The expansion was completed under budget and ahead of schedule, both unusual on projects of this size and complexity. What is more, the plant is of high quality, incorporating all the improvements normally omitted by an outside designer. Work was done by local contractors using equipment predominently of Canadian manufacture.

As in any project, there was a lot of hard work and many frustrations. But the co-operation and goodwill of the many people concerned working together as a team are the main ingredients for success in such a venture. And the result is a project which makes us all very proud.

Discussion of Paper by Tony Hill

R. Tenbergen:	Did you consider autogenous grinding?
Reply:	Originally, they considered semi and fully autogenous grinding but because of a lack of time and experience, we stayed with the conventional grinding.
B. Ferguson:	Who were the original contractors ?
Reply:	Ralph M. Parsons.

Paper No. 4

LABORATORY INVESTIGATIONS AND IMAGE ANALYSIS STUDIES ON SAMPLES FROM THE BASE METAL DEPOSIT OF BRUNSWICK MINING AND SMELTING CORP. LTD.

by

A.I. Stemerowicz* and W. Petruk*

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ABSTRACT

Laboratory investigations were conducted on samples from the base metal deposit of Brunswick Mining and Smelting in New Brunswick to develop flotation techniques for recovering zinc from mill tailings and for producing bulk concentrates with high metal recoveries from the ore. Test products from the laboratory investigations were studied by image analysis to evaluate the behaviour of the minerals during flotation. A mill tailing grading 1.8 wt % zinc was reground to 80% minus 19 micrometers and floated using conventional flotation methods. A zinc concentrate grading 45.5 wt % zinc was produced with a recovery of 62%. Sphalerite liberation in the reground mill tailing was 78% and recovery in the zinc concentrate was 70% of the free and 35% of the unliberated sphalerite. An ore grinding 8.67 wt % zinc was ground to 77.5% minus 500 mesh and treated by conventional flotation techniques to produce a bulk concentrate grading 30 wt % zinc and containing 95.3% of the zinc. Sphalerite liberation in the ground ore was 86% and recovery was 98% of the free sphalerite and 84% of the unliberated sphalerite.

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INTRODUCTION

The laboratories in the Mineral Sciences Division at CANMET, Department of Energy, Mines and Resources are involved in a program that has the objective of establishing methods for improving metal recoveries from the base metal ores of New Brunswick. One phase of the program was to develop a flotation technique for recovering zinc from mill tailings, and another was to develop a flotation technique for producing bulk concentrates from the ores with high metal recoveries. The bulk concentrates would be used for subsequent hydrometallurgical and pyrometallurgical treatments. The investigations were conducted by conventional batch flotation methods, and the resulting mill products were studied by cyclosizing, chemical and/or XRF analysis, and image analysis. This paper is presented to indicate the recoveries of metals, of free sphalerite, of slimed sphalerite, and of unliberated sphalerite that were obtained in these investigations.

RECOVERING ZINC FROM MILL TAILINGS

The mill of Brunswick Mining and Smelting Corporation Limited (BMS) produces a primary mill tailing that contains about 15% of the zinc in the ore and has a grade of 1.8 wt % zinc. The primary mill tailing is treated in a tailings flotation circuit that recovers about 35% of the zinc and produces a final mill tailing that contains about 10% of the zinc in the ore and has a grade of 1.2 wt % zinc. Samples of the primary mill tailing were investigated at CANMET to establish a flotation technique for recovering more of the zinc than the company currently recovers.

Preliminary batch flotation tests showed that in order to improve metal recoveries it would be necessary to regrind the material. The grind of the primary mill tailing, as received, was about 80% minus 270 mesh. About 39% of the contained sphalerite occurred as free grains with about one-half of these being smaller than 6 μm(1). After regrinding for one hour in a laboratory mill the grind was about 80% minus 19 micrometers; 78% of the sphalerite was free with about one-half of the free sphalerite being smaller than 6 μm. A concentrate grading 45.5 wt % zinc was obtained, with a recovery of 62%, by utilizing one rougher stage and cleaning five times in open circuit. The calculated grade of the laboratory mill tailing is 0.72 wt % zinc and the calculated zinc loss in this tailing is 6% of the zinc in the BMS ore. Image analysis studies showed that 70% of the free sphalerite and 35% of the unliberated sphalerite were recovered in the concentrates. The unliberated sphalerite (middlings) in the concentrate was in composite grains that contained an average of 28 wt % sphalerite. The average surface exposure of the sphalerite in these grains was 70%(2).

Size analyses were performed with the image analyser of the free and unliberated sphalerite in each product. The size of unliberated sphalerite was taken as the size of sphalerite within the composite grain. The size analysis data were converted to absolute quantities of free and unliberated sphalerite on the basis of total sphalerite in the test feed being 100%. The results were then plotted as frequency diagrams of absolute quantity vs size for the free and unliberated sphalerite in the feed and the rougher and cleaner concentrates (Fig. 1). The difference between points on the curve for the feed and for the rougher concentrate is equal to the absolute quantity of sphalerite lost in the rougher tailings. Similarly, the difference between points on each successive curve for the cleaner concentrates is equal to the sphalerite lost in each successive tailing. The top curve (feed) shows that most of

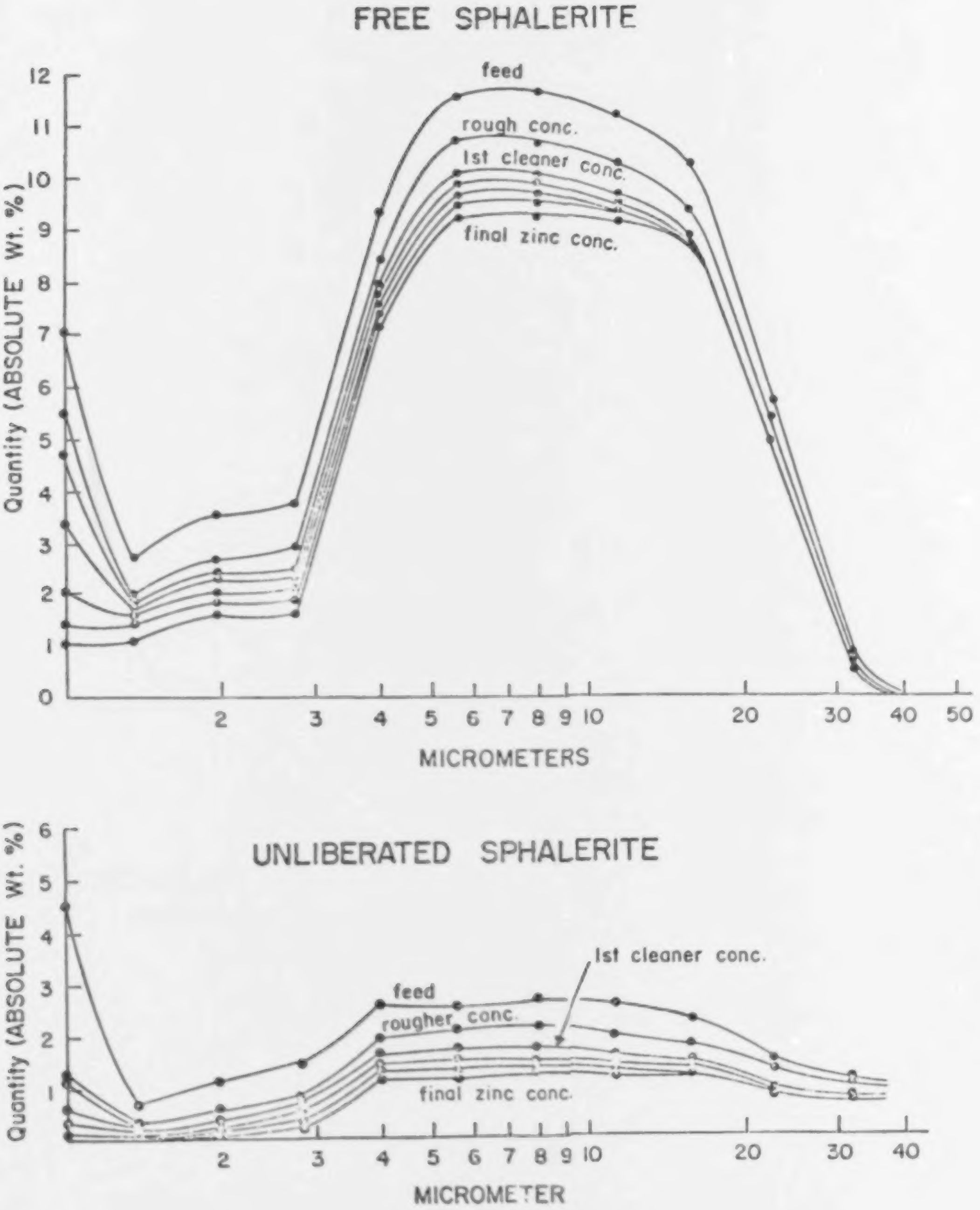
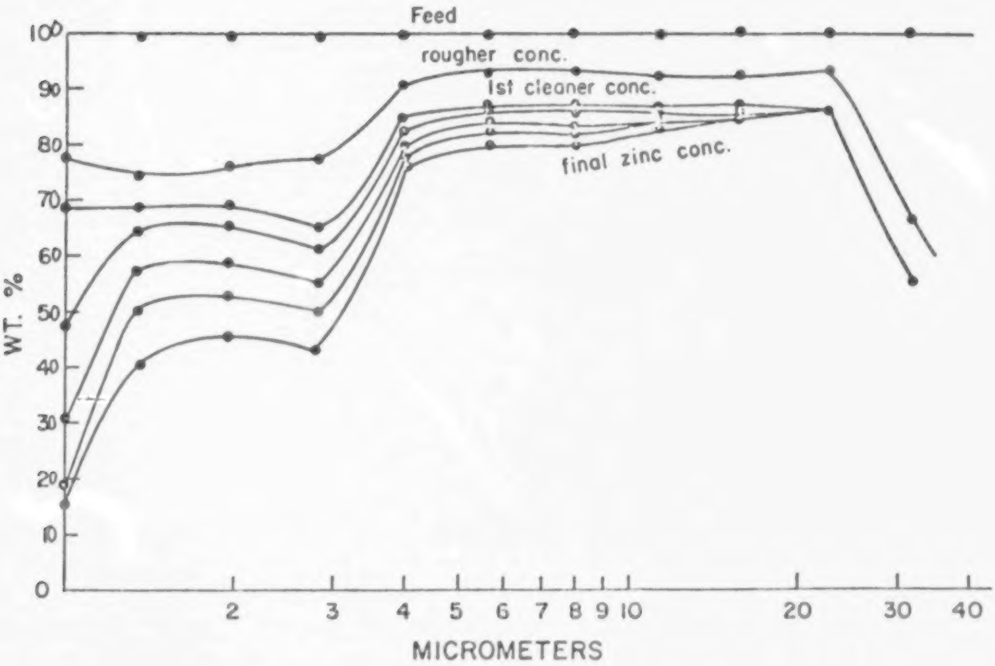


Figure 1. Frequency diagrams showing absolute quantities of sphalerite vs size.

FREE SPHALERITE



UNLIBERATED SPHALERITE

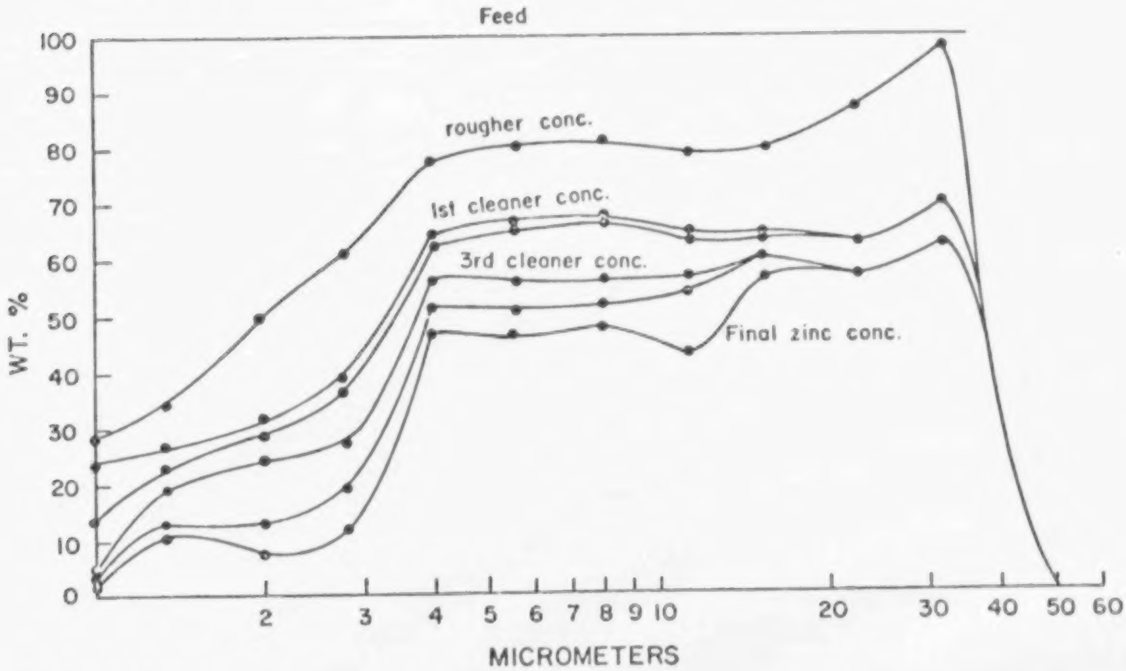


Figure 2. Frequency diagrams for sphalerite on basis of wt % at each size interval.

the free and unliberated sphalerite grains in the feed were 3 to 25 μm in diameter and minus 1 μm, respectively. The curve for the final zinc concentrate shows that most of the minus 1 μm, and indeed much of the minus 3 μm free sphalerite was lost in the rougher plus cleaner tailings.

To further evaluate the sphalerite recovery, the quantity of both free and unliberated sphalerite in the feed was made to 100% for each size range and the percentage removed in the successive tailings was subtracted from the 100% value to give values for sphalerite in the successive concentrates. These were plotted as frequency diagrams (Fig. 2). The curve for sphalerite in the final zinc concentrate shows that recovery of free sphalerite was 85% for grains 13 to 26 μm in diameter; recovery dropped linearly to 75% for the grains 3.6 to 4.6 μm in diameter; dropped sharply to 42% for 1.2 to 3.6 μm grains; and was 15% for minus 1.2 μm grains. A similar evaluation shows a lower recovery of unliberated sphalerite (Fig. 2) with nearly all the minus 3 μm unliberated sphalerite having been discarded.

The curves in Figure 2 show that free sphalerite grains of all sizes were lost in the rougher and first cleaner tailings. On the other hand, the free sphalerite lost in the second to fifth cleaner tailings was negligible in the coarse sizes but increased below 4 μm.

PRODUCING A BULK CONCENTRATE

Object and Overall Scope of Investigation

This investigation was undertaken to develop flotation techniques for producing a bulk concentrate that would contain about 30 wt % zinc, with optimum recoveries of zinc, lead, copper and silver. Ore from the BMS deposit was used for the tests and it had a grade of 8.76 wt % zinc, 3.42 wt % lead, 0.25 wt % copper, and 2.3 oz/ton silver.

Bulk concentrates were produced by direct bulk flotation of the valuable minerals and by selectively floating separate lead and zinc concentrates and then combining the two concentrates and some of the cleaner tailings.

The flotation variables investigated were (1) fineness of grind for both bulk and selective flotation, (2) high versus low pH in bulk roughers and cleaners, (3) coarse primary grind for selective flotation in combination with regrinding of lead and zinc rougher concentrates prior to cleaning, (4) gypsum versus lime as a pyrite depressant during bulk flotation, and (5) two levels of copper sulphate addition to bulk flotation. The separation efficiency as expressed by Schultz(4) was used to compare the tests because the recoveries of several metals were involved. This separation efficiency was calculated by subtracting the percent recovery of unwanted gangue minerals in the concentrate from the percent recovery of valuable minerals concentrated. The quantities and recoveries of minerals were calculated from the chemical analysis on the assumption that the valuable minerals, sphalerite, galena and chalcopyrite, contain 60 wt % zinc, 86.6 wt % lead, and 34.5 wt % copper, respectively. The remainder was assumed to be unwanted gangue minerals. The concentrates and tailings from some of the tests were studied by image analysis(5).

Selective Flotation Versus Direct Bulk Flotation

Tests were conducted to determine whether better results could be obtained by selective flotation or by direct bulk flotation to produce a bulk concentrate with a target grade of 30 wt % zinc. In the case of selective flotation a lead and a zinc concentrate were produced according to the flowsheet in Figure 3 and the two concentrates and some of the zinc cleaner tailings were combined to make the target grade concentrate. For direct bulk flotation a bulk cleaner concentrate was produced (Fig. 4) and combined with an appropriate amount of cleaner tailings to make the target concentrate grade. A comparison of the metal recoveries and separation efficiencies for concentrates produced by the two methods shows that better results were obtained by selective flotation (Table 1).

TABLE 1

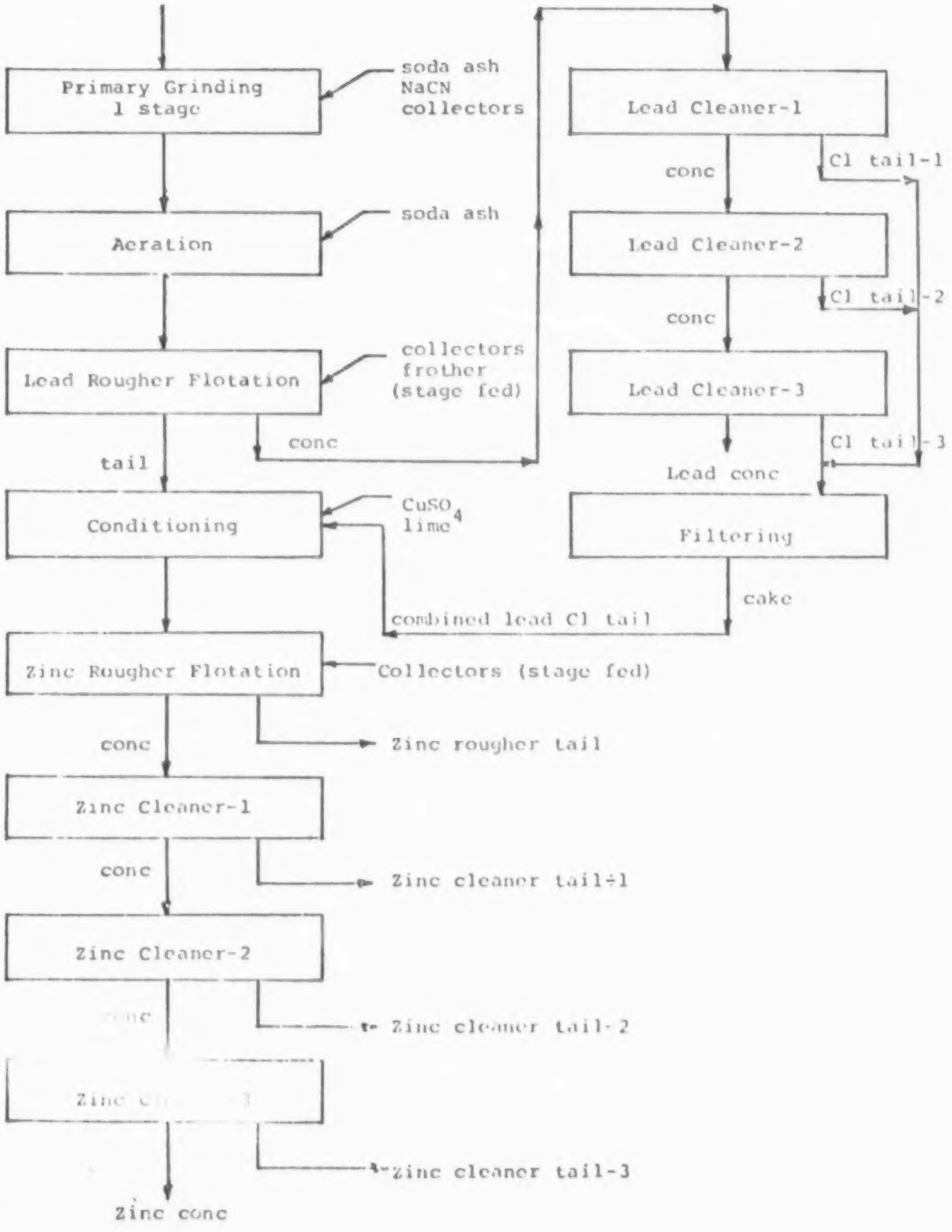
Comparison of Bulk Concentrate Produced by Selective Versus Direct Bulk* Flotation

Test No.	Flotation Method	Wt %	Analysis (Wt %)			Recoveries %			Sep. Eff.
			Zn	Pb	Cu	Zn	Pb	Cu	
A-21	selective	27.54	30	10.8	0.7	95.3	86.6	76.7	80.7
A-30	direct bulk	27.15	30	11.2	0.7	92.2	81.9	74.1	77.7

*Grind = 77.5% minus 500 mesh.

Effect of Fineness of Grind

Table 2 gives a comparison of results obtained by selective flotation and by direct bulk flotation of the ore ground to two different degrees of fineness. Somewhat higher recoveries and separation efficiencies were obtained for the finely ground feed.


Primary Grinding
1 stage
soda ash
NaCN
collectors
Aeration
soda ash
Lead Rougher Flotation
collectors
frother
(stage fed)
conc
tail
Conditioning
CuSO4
lime
Lead Cleaner-1
conc
Cl tail-1
Lead Cleaner-2
conc
Cl tail-2
Lead Cleaner-3
Cl tail-3
Lead conc
Filtering
cake
combined lead Cl tail
Zinc Rougher Flotation
Collectors (stage fed)
conc
Zinc rougher tail
Zinc Cleaner-1
conc
Zinc cleaner tail-1
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Zinc cleaner tail-3
Zinc conc

Figure 3. Test flowsheet for selective flotation.

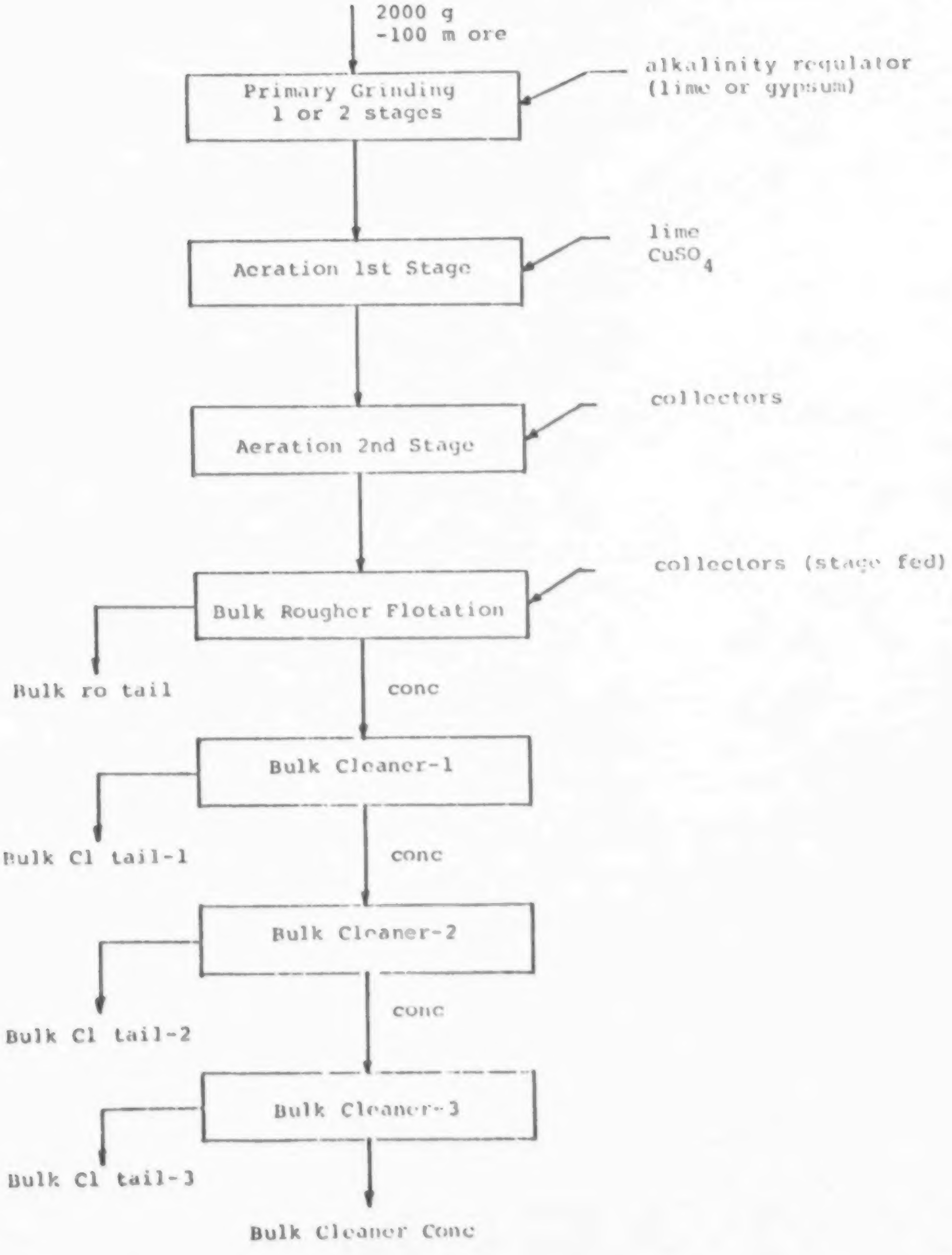
2000 g
-100 m ore
Primary Grinding 1 or 2 stages
alkalinity regulator (lime or gypsum)
Aeration 1st Stage
lime
CuSO4
Aeration 2nd Stage
collectors
Bulk Rougher Flotation
collectors (stage fed)
Bulk ro tail
conc
Bulk Cleaner-1
Bulk Cl tail-1
conc
Bulk Cleaner-2
Bulk Cl tail-2
conc
Bulk Cleaner-3
Bulk Cl tail-3
Bulk Cleaner Conc

TABLE 2

Comparison of Effect of Fineness of Grind

Test No.	Flotation Method	Grind % - 500 Mesh	Wt %	Analysis (Wt %)			Recoveries %			Sep. Eff.
				Zn	Pb	Cu	Zn	Pb	Cu	
A-21	selective	77.5	27.54	30.0	10.8	0.7	95.3	86.6	76.7	80.7
A-23	selective	57.0	27.74	30.0	11.4	0.7	94.3	86.2	71.9	79.5
A-30	direct bulk	77.5	27.15	30.0	11.2	0.7	92.2	81.9	74.1	77.7
A-27	direct bulk	57.0	25.79	29.5	11.6	0.7	87.9	80.1	67.4	73.9

Effect of pH Level in Bulk Rougher Flotation and Cleaners

Tests were conducted at high and low pH in the roughers and cleaners. The ranges of pH values employed were:

roughers: high pH, 11.4 at start to 10.3 at end of float.
low pH, 9.8 at start to 9.0 at end of float.

cleaners: high pH, 11^+ in each cleaning stage.
low pH, 9.9 in first cleaner and 8.6 in final cleaner.

Better selectivity accompanied by lower zinc and lead tailing losses was obtained in tests employing a high pH in the roughers (Table 3) and a low pH in the cleaners (Fig. 5).

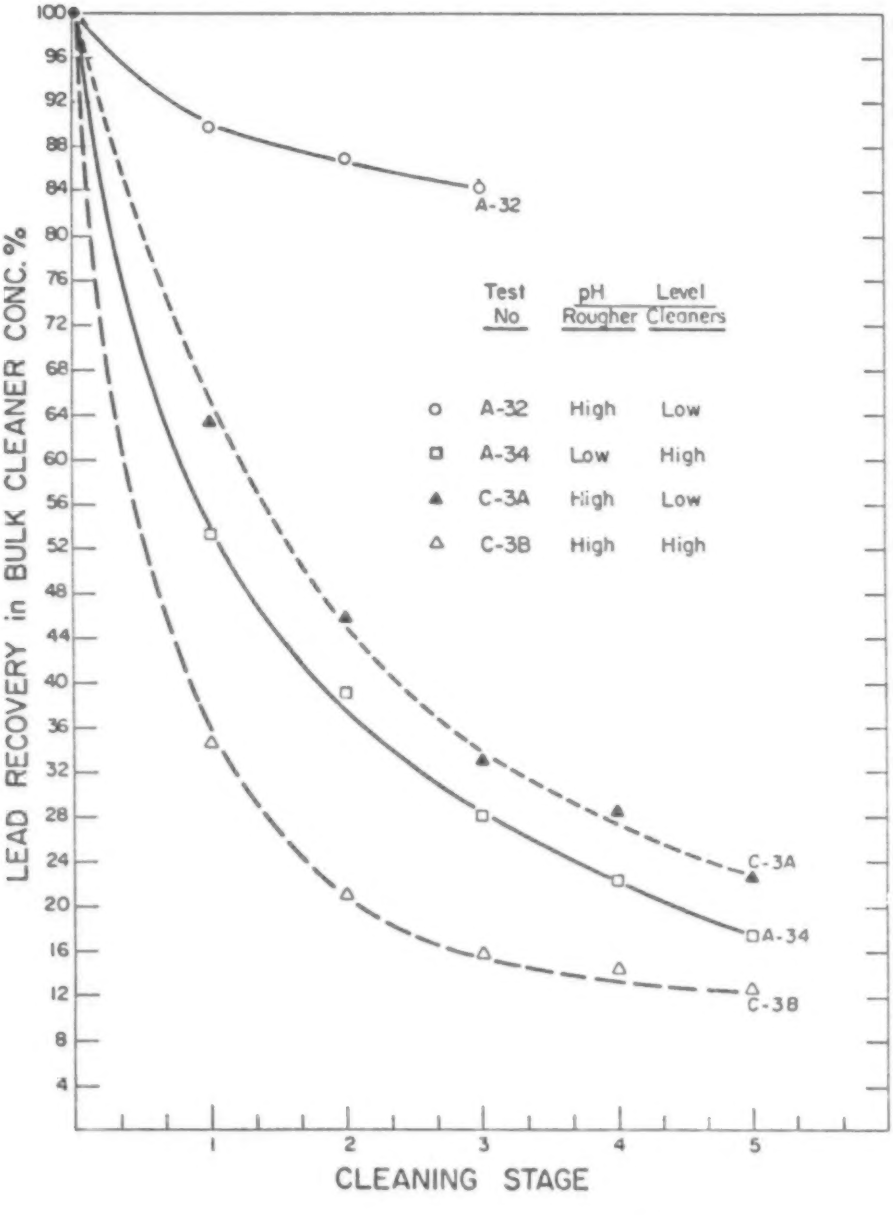

Figure 5. Lead recovery at different cleaning stages using high and low pH in cleaners. Tests A-32 and A-34 were conducted on one sample and tests C-3A and C-3B on another sample.

TABLE 3

Results for Rougher Flotation Concentrates at Low and High pH

Test Level	pH Level	Wt %	Assay (Wt %)			Distribution %			Sep. Eff.
			Zn	Pb	Cu	Zn	Pb	Cu	
A-30	low	46.62	18.4	7.2	0.5	97.0	91.3	86.0	60.9
A-31	high	43.80	19.7	7.4	0.5	97.9	91.4	85.7	64.8
A-32	high	43.32	19.3	7.3	0.5	97.3	90.7	87.4	63.9
A-34	low	60.57	14.1	5.2	0.4	97.2	92.5	88.0	49.7

Regrinding Prior to Cleaning

A test was conducted to compare the recovery of metals by selective flotation from ore that had a fine primary grind with the recovery of metals from ore that had a coarse primary grind but the lead and zinc rougher concentrates were reground prior to cleaning. Essentially the same results were obtained for material ground by the two procedures (Table 4), except that the copper recovery was lower for the coarse primary grind.

TABLE 4

Comparison of Fine Primary Grind with Coarse Primary Grind and Regrinding Prior to Cleaning

Test No.	Grind % -500 mesh	Wt %	Analysis (Wt %)			Recoveries %			Sep. Eff.
			Zn	Pb	Cu	Zn	Pb	Cu	
A-21	Primary 77.5	27.54	30	10.8	0.7	95.3	86.6	76.6	80.7
A-25	Primary 57.0, Reground	24.84	30	11.1	0.7	95.2	85.8	67.9	81.2

Gypsum Versus Lime During Grinding

Table 5 gives a comparison of bulk flotation results for two pairs of tests (A-7, A-9 and A-15, A-17) that were designed to test the effectiveness of gypsum as a pyrite depressant in place of lime during grinding. The results show that gypsum did not exhibit any particular advantage over lime.

TABLE 5

Comparison of Gypsum Versus Lime During Grinding

Test No.	Material Added	pH	Wt %	Analysis (Wt %)			Recoveries %			Sep. Eff.
				Zn	Pb	Cu	Zn	Pb	Cu	
A-7	gypsum	9.0	35.21	23.1	9.3	0.6	92.9	87.5	78.6	69.5
A-9	lime	10.0	36.28	22.6	9.4	0.6	93.4	88.6	83.1	69.4
A-17	gypsum	9.0	35.15	23.2	8.7	0.6	93.8	83.0	78.7	69.1
A-15	lime	9.8	30.70	22.7	9.6	0.7	93.5	87.3	80.7	73.3

Effect of Copper Sulphate Addition

The standard amount of copper sulphate added for sphalerite activation in both the direct bulk and selective flotation was 2.0 lb/ton. In some tests by direct bulk flotation 3.0 lb/ton copper sulphate was added. Table 6 shows that the increase in copper sulphate addition resulted in an increased zinc recovery without adversely affecting selectivity.

TABLE 6

Effect of Copper Sulphate Level in Direct Bulk Flotation

Test No.	Grind % - 500 Mesh	$CuSO_4$ lb/ton	Wt %	Analysis (Wt %)			Recoveries %			Sep. Eff.
				Zn	Pb	Cu	Zn	Pb	Cu	
A-28	77.5	2.0	46.83	15.0	6.5	0.4	92.4	90.0	79.1	53.7
A-30	77.5	3.0	46.62	18.4	7.2	0.5	97.0	91.3	86.1	60.6
C-1	57.0	2.0	27.60	30.1	8.1	0.6	88.7	62.4	69.6	69.3
C-2	57.0	3.0	30.38	29.6	8.0	0.6	94.0	69.6	70.6	73.2

Image Analysis Study of Test Products

Image analysis studies were conducted on the concentrates and tailings from the four tests reported in Table 2. The results show that more sphalerite and galena were liberated in the finely ground material than in the coarsely ground ores (Table 7). The proportions of both free and unliberated sphalerite recovered by selective flotation were higher than by direct bulk flotation (Table 8), hence the better zinc recovery by selective flotation. It is noted that by selective flotation the same percentage free sphalerite was recovered from both the finely ground and coarsely ground feeds, but a higher percentage of the unliberated sphalerite was recovered from the coarsely ground feed. By direct flotation the same percentage of unliberated sphalerite was recovered from both the coarsely ground and finely ground feeds, but a much lower percentage of free sphalerite was recovered from the coarsely ground feed than from the finely ground feed. This low recovery of free sphalerite is anomalous which suggests that a higher zinc recovery could have been obtained in Test A-27.

TABLE 7

Liberation and Recovery of Metallic Minerals

Sample No.	Grind % - 500 Mesh	% Free		Recovery of sphalerite (Absolute Wt %)		
		Sphalerite	Galena	as free grains	as unlib. grains	Total
A-21	77.5	86	60	82.0	13.3	95.3
A-23	57.0	66	42	62.3	32.0	94.3
A-30	77.5	84	56	77.5	14.8	92.3
A-27	57.0	70	41	61.6	26.3	87.9

TABLE 8

Effect of Flotation Method on Recovery of Free and Unliberated Sphalerite

Sample No.	Flotation Method	Grind % - 500 Mesh	% of Free Sphalerite Recovered	% of Unliberated Sphalerite Recovered
A-21	selective	77.5	98.1	83.8
A-23	selective	57	98.1	87.4
A-30	direct bulk	77.5	97.4	75.6
A-27	direct bulk	57	92.4	75.3

Size analyses were performed on the concentrates and tailings and the absolute quantities of sphalerite within each size range recovered in the concentrates and lost in the tailings were calculated. The results for absolute quantities of sphalerite lost in the tailings are given in Table 9. This table shows that, for all tests, the highest loss of free sphalerite was as grains smaller than 4.7 μm. On the other hand, unliberated sphalerite grains of all sizes were lost. The concentrates and tailings from these tests were cyclosized and the sized fractions were analysed. The results, summarized in Table 10, show that a high proportion of the metals was lost in the minus 4.7 μm fractions.

TABLE 9

Size Analyses of Free and Unliberated Sphalerite Lost in Tailings

Sample No.	Flotation Method	Grind % - 500 Mesh	Sphalerite loss (Absolute Wt %)							
			Free Sphalerite				Unliberated Sphalerite			
			+26 µm	26-4.7 µm	-4.7 µm	Total	+26 µm	26-4.7 µm	-4.7 µm	Total
A-21	selective	77.5	0	0.3	1.3	1.6	0	0.6	2.0	2.6
A-23	selective	57	0	0.3	0.9	1.2	1.3	1.3	2.0	4.6
A-30	direct bulk	77.5	0	0.4	1.7	2.1	0.9	1.0	2.9	4.8
A-27	direct bulk	57	0	1.4	3.6	5.0	1.8	2.1	4.6	8.5

TABLE 10

Size Analysis of Zinc Lost in Tailings
(Absolute Wt %)

Sample No.	Flotation Method	Grind % - 500 Mesh	Zinc Lost (Absolute Wt %)			
			+26 µm	26-4.7 µm	-4.7 µm	Total
A-21	selective	77.5	0.7	2.0	2.0	4.7
A-23	selective	57.0	2.1	1.5	2.1	5.7
A-30	direct bulk	77.5	0.7	2.3	4.8	7.8
A-27	direct bulk	57.0	3.5	2.7	5.9	12.1

Relationship Between Zinc Recoveries and Sphalerite in Mill Feed

A general relationship between zinc recoveries and the characteristics of sphalerite in a mill feed can be deduced by comparing the results from the investigation on mill tailings with the results from the investigation on producing a bulk concentrate (Table 11). The comparison is considered to be valid because both investigations were conducted on material from the same ore hence the same ore and gangue minerals were present in the flotation feeds. Furthermore the sizes of the free sphalerite grains were similar (Fig. 6). The data in Table 11 suggest that sphalerite recoveries were directly proportional to sphalerite liberation and feed grade, and inversely proportional to concentrate grade, concentration ratios, quantities of minus 4.7 μm free sphalerite in feed and pyrite:sphalerite ratios in feed.

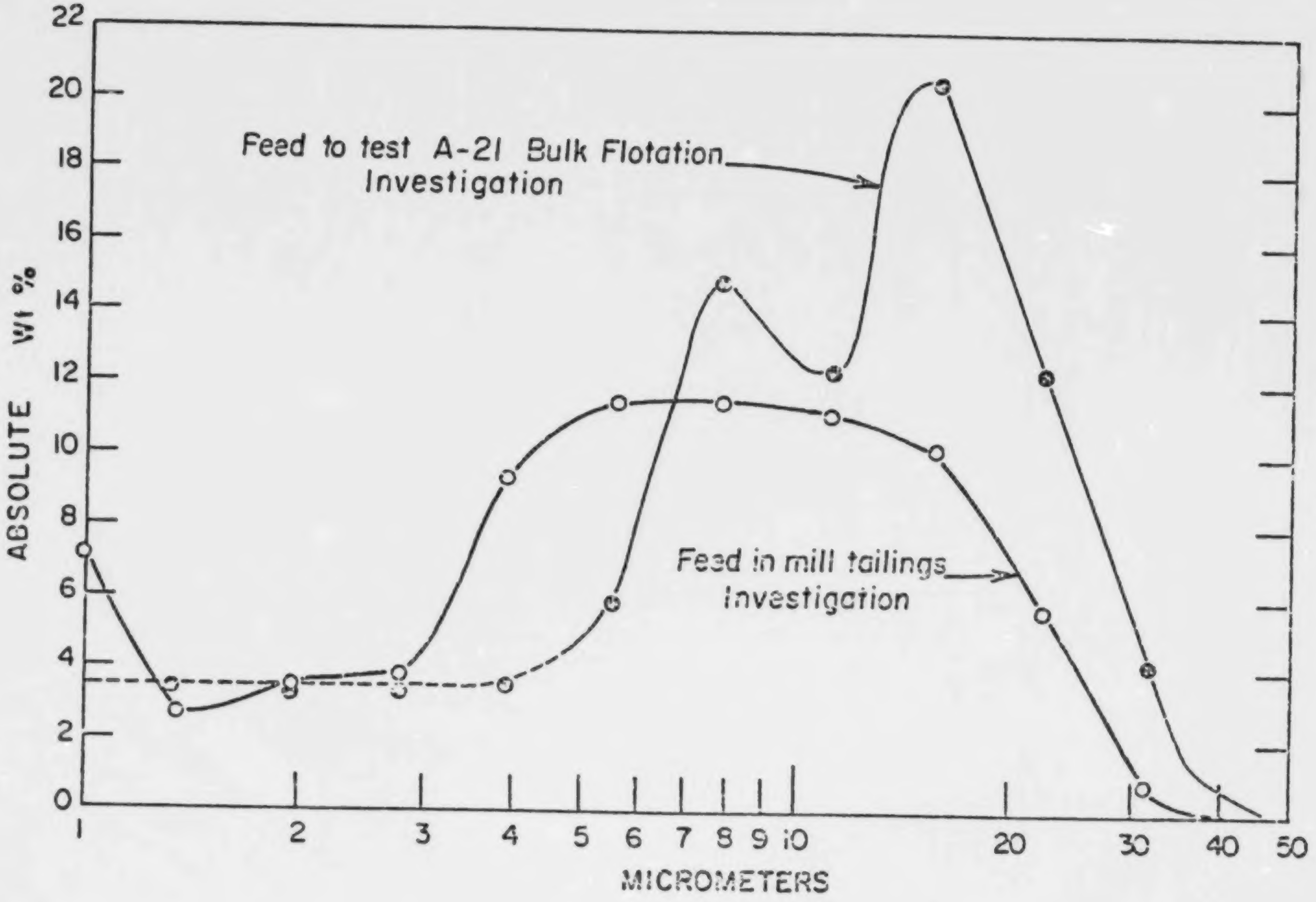

Figure 6. Frequency diagrams showing absolute quantities of free sphalerite vs size in the feed to mill tailings investigation and in the feed to the bulk flotation investigation.

TABLE 11

Comparison of Sphalerite in Mill Tailing and Bulk Concentrate Investigations

	Mill Tailing	Bulk Conc.
Feed grade (wt % Zn)	1.8	8.7
Concentrate grade (wt % Zn)	45.5	30.0
Concentration ratio $\frac{\text{wt of feed}}{\text{wt of conc.}}$	39.5	3.6
Zinc recovery (%)	62.0	95.3
Sphalerite liberation (% as free grains)	78	86
% of sphalerite as free grains at -4.7 μm	34	16
Size range of most of free sphalerite* (μm)	3-30	5-37
Sphalerite recovery (1) % of +4.7 μm free grains	82.2	99.7
(2) % of -4.7 μm free grains	46.7	90.3
(3) % of unliberated	35.0	84.0
Pyrite:Sphalerite ratio in feed	20:1	4:1

*From Fig. 6.

REFERENCES

1. Petruk, W. and Pinard, R.G. "Image analysis studies of tailings from the mill of Brunswick Mining and Smelting Corporation in New Brunswick"; CANMET, Energy, Mines and Resources Canada; *Lab*. *Report* MRP/MSL 77-8 (IR); 1977.

2. Petruk, W. "Image analysis study of mill products from batch tests of Brunswick Mining and Smelting mill tailings"; CANMET, Energy, Mines and Resources Canada; *Lab*. *Report* MRP/MSL 78-187 (R); 1978.

3. Stemerowicz, A.I. and Leigh, G.W. "Flotation techniques for production of high-recovery bulk Zn-Pb-Cu-Ag concentrate from a New Brunswick massive sulphide ore"; CANMET, Energy, Mines and Resources Canada; *Lab*. *Report* MRP/MSL 78-174 (R); 1978.

4. Schultz, N.F. "Separation efficiency"; *Soc*. *Min*. *Eng*. *AIME* *Transactions* 247; 81-87; 1970.

5. Petruk, W. "Image analysis of bulk concentrates and tailings produced in the Ore Processing Laboratory from Brunswick No. 12 ore"; CANMET, Energy, Mines and Resources Canada; *Lab*. *Report* MRP/MSL 78-188 (IR); 1978.

Discussion of Paper by A. Stemerowicz and W. Petruk

G. Johnston: In your analysis of the minerals, were you dealing with sphalerite of marmatitic minerals ?

Reply: The sphalerite in the ore varies from about 4% to 10% Fe and averages about 6%. I could not distuish the various ore minerals. The average is about 60% Zn and 40% Fe.

K. Konigsmann: How about silver in the ore ?

Reply: The silver is associated with tetrahedrite and galena. We know that it is always with the sulphides and in a bulk concentrate we get a recovery of 85% of the silver. The silver loss in the tailing is due to the fine grained galena and copper minerals. If we can recover the fine grained galena and copper minerals, we can recover the silver as well.

Paper No. 5

MULTI-STAGE HYDROCYCLONE CIRCUIT OPTIMIZATION BY COMPUTER SIMULATION

R. STRATTON-CRAWLEY* AND G.E. AGAR**

ABSTRACT

The applicability of a hydrocyclone model to simulation of the cyclone banks in a multi-stage sandplant circuit producing mine back fill has been demonstrated. The model has been used in conjunction with a modular computer program for circuit simulation in order to study the effect of changes in the operating conditions. It has been shown that weight recovery to sand fill increases as feed percent solids decreases, and, to a lesser extent as pressure drop across the banks increases. An increase in apex diameter also causes an increase in weight recovery. Increased weight recovery must, however, be balanced against sump and cyclone capacity and dilution of sand fill by fines. The predicted results from circuit simulation were in good agreement with plant data.

The model has been found to be applicable to widely different ores.

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INTRODUCTION

Flotation tailings at the INCO Thompson, Manitoba mill, are treated in a sandplant circuit to produce sandfill for the Thompson mine. The quantity of sandfill produced (approximately 30% of the Thompson mill feed) has not been sufficient to meet backfill requirements, the amount of fill necessary to sustain the mining cycle being 52% (1). The remainder of the fill requirement is obtained from Pipe Open Pit tailings and is occasionally supplemented by slag and alluvial sand. The failure of the sandplant to supply backfill needs was due, in part, to old and worn cyclones in the circuit which performed poorly. This situation led to the installation of 8 Krebs D10LB-845 cyclones in the primary bank. Several sampling campaigns were then carried out around these cyclones to evaluate their behaviour and determine their effect on circuit performance. The results obtained from these surveys were used to develop a mathematical model of the hydrocyclone which would accurately describe cyclone behaviour and be suitable for incorporation into a computer program which could simulate the entire circuit under a wide range of operating conditions. Subsequently, upon installation of new cyclones in the cleaner bank, the computer program was used to optimize the operating conditions in the new circuit.

THE MODEL

Mathematical models describing hydrocyclone operation may be divided into analytical and empirical. The complex flow regimes with in a cyclone have generally meant that analytical models are unsuitable for practical simulation purposes (2). There are, however, a large number of satisfactory empirical models which describe cyclone behaviour with varying degrees of success. Much of the early work utilized data from small cyclones obtained with dilute slurries as the experimental basis for the models, for this reason most are not applicable to the majority of mineral processing operations. In recent years, however, several workers have studied the performance of large hydrocyclones operating with high feed percent solids, notable amongst these is the work of Lynch and Rao (2) and Plitt (3). Lynch and Rao developed a number of empirical equations describing cyclone performance, but the applicability of the model was limited as many of the design and operating variables (such as hydrocyclone diameter and feed size distribution) were kept constant. Plitt, however, in an exhaustive investigation developed a more universally applicable model. Indeed, the model was found to be in general agreement with a number of analytical models.

In the present investigation, Plitt's model was taken for the basis of the computer program and modified in those areas where it did not describe plant cyclone behaviour sufficiently accurately. The model consists of seven basic equations which define cyclone capacity, volume split of water between underflow and overflow, cut size, probability of particles of given size reporting to the underflow, separation modulus, cyclone separation efficiency and a correction factor describing classification behaviour of -45 μm particles. These parameters are related to the operating and design variables of the cyclone and are summarized below:

(i) Capacity equation

The relationship between the volumetric flow-rate and the pressure drop across the cyclone is given by

$$Q = \frac{P^{0.56}\ D_c{}^{0.21}\ D_i{}^{0.53}\ h^{0.16}\ (D_u{}^2 + D_o{}^2)^{0.49}}{K_p \exp\left[0.0031\phi\right]} \qquad \ldots\ldots\ldots \quad 1$$

The nomenclature used in this and subsequent equations is given following the text.

(ii) Volume Split Equation

The amount of water that flows to the underflow is a measure of the feed solids that by-pass the cyclone without undergoing classification. The volume split, S, between underflow and overflow defines this factor.

$$S = \frac{\text{U/F rate}}{\text{O/F rate}} = \frac{k_s(D_u/D_o)^{3.31}h^{0.54}(D_u^{\,2}+D_o^{\,2})^{0.36}\exp[0.0054\phi]}{P^{0.24}D_c^{\,1.11}} \qquad \ldots . \quad 2$$

In addition, the flow ratio, R_f, is given by

$$R_f = \frac{\text{U/F rate}}{\text{Feed rate}} = \frac{S}{S+1} \qquad \ldots . \quad 3$$

Equation 2 was derived for cyclones operating with free discharge and is most sensitive to back pressure and syphoning effects.

(iii) Cut Size Equation

The separation size ($d_{50(c)}$) of the cyclone is one of the major parameters determining cyclone performance and is given by

$$d_{50(c)} = \frac{K_D D_C^{\,0.46} D_i^{\,0.6} D_o^{\,1.21}\exp\,[0.063\phi]}{D_u^{\,0.71}h^{0.38}Q^{0.45}(\rho_s - \rho)^{0.5}} \qquad \ldots . \quad 4$$

(iv) Reduced Efficiency Equation

The reduced efficiency curve defines the probability of a particle of given size reporting to the cyclone underflow. The curve has been found to fit the Rosin-Rammler-Bennet equation of probability (4), viz..

$$E_{Net} = 1 - \exp\left[-0.693(d/d_{50(c)})^{m}\right] \quad \ldots\ldots \quad 5$$

where m is given by

$$\text{(v)} \quad m = K_m\left(\frac{D_c{}^2 h}{Q}\right)^{0.15} \quad \ldots\ldots \quad 6$$

The above equations were all reported to have correlation coefficients equal to or greater than 0.9(3) except for equation 6 for which C=0.75, for cyclone sizes ranging from 1 1/4" - diameter to 20" - diameter. The data for the larger cyclones was obtained from the work of Lynch and Rao (2).

In addition to these the following equation was used to define separation efficiency (5),

$$\eta = G_u\left(\frac{1-(Y_{d50})_u}{1-(Y_{d50})_f} - \frac{(Y_{d50})_u}{(Y_{d50})_f}\right) \quad \ldots\ldots \quad 7$$

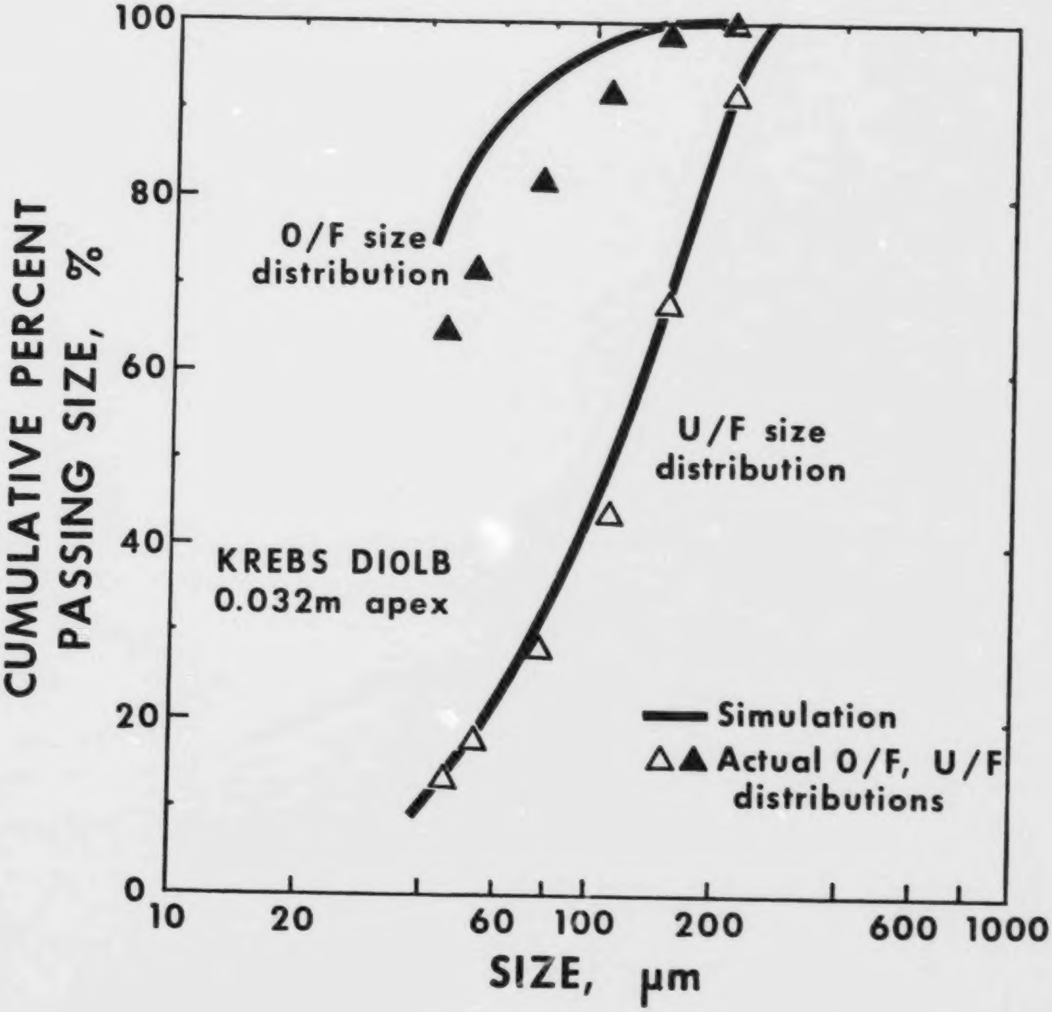
where, G_u is $\dfrac{\text{mass flow-rate of solids in underflow}}{\text{mass flow-rate of solids in feed}}$

and Y_{d50} is the cumulative fraction of solids finer than d_{50} in the designated stream.

It was also found necessary to include an equation defining the probability of -45 μm particles reporting to the cyclone underflow as equation 5 was inaccurate for the fine size fraction. This derivation of this equation is described below.

Figure 1

COMPARISON OF ACTUAL AND SIMULATED SIZE DISTRIBUTION USING ORIGINAL MODEL CONSTANTS



DATA PROCESSING

Upon installation of the Krebs DIOLB cyclones in the primary cyclone bank treating tailings from the circuit processing Thompson Mine are (hereafter referred to as B circuit), a number of sampling surveys were carried out around the bank. Table 2 gives a typical set of results for cyclones operating under the conditions shown in Table 1. Initially the behaviour of the hydrocyclone operating under these conditions was simulated using the above model and the constants recommended by Plitt (3). Figure 1 shows the product size analyses obtained from the simulation, these are compared to the distributions observed in practice. In the case of the underflow

TABLE 1: DESIGN AND OPERATING VARIABLES OF KREBS D1OLB-845 CYCLONES INSTALLED AS B SANDPLANT PRIMARIES

Cyclone dimensions		
	Cylinder internal diameter	0.254 m
	Inlet diameter	0.102 m
	Vortex finder diameter	0.089 m
	Free vortex height	0.787 m
	Apex diameter	0.032 m
	Cone angle	22°
Operating conditions		
	Feed % solids	29%
	Pressure drop	125 kPa

TABLE 2: 'B' SANDPLANT PRIMARY CYCLONES-MASS BALANCE AND STREAM SIZE DISTRIBUTIONS

Stream	% Solids	Solids % Wt Recovery	Cumulative Weight % on Screen Size (µm)						
			+212	+150	+106	+ 75	+ 53	+ 45	- 45
Feed	29.2	100.0	4.9	16.6	30.7	43.7	53.1	59.2	40.6
O/F	18.9	53.6	0.6	2.3	8.0	18.3	27.9	35.4	64.6
U/F	67.7	46.1	9.9	33.3	57.3	73.4	82.7	87.2	12.9
			% Weight Recovery of Screen Size Fraction						
			+212	+150	+106	+ 75	+ 53	+ 45	- 45
Feed			100.0	100.0	100.0	100.0	100.0	100.0	100.0
O/F			6.4	8.0	21.5	42.7	54.7	66.2	85.4
U/F			93.6	92.0	78.5	57.3	45.3	33.8	14.6

there is good agreement between the actual and simulated size distribution. The predicted overflow size distribution is significantly finer than that obtained in the plant however. Some additional data is compared in Table 3. There is general agreement between most of the simulated and actual results, except in the case of the pressure drop required to achieve the actual volume flow-rate, where the predicted pressure is significantly lower than that actually required. The simulated distribution of solids between underflow and overflow is also not sufficiently accurate, however, for immediate application of the model for practical simulation purposes. The predicted recovery of solids to the sandfill product using the conditions given in Table 1 is 58.8%, 12.7% higher than that found in practice, clearly an overoptimistic evaluation.

Figure 2

PERFORMANCE CURVES FOR KREBS DIOLB CYCLONES WITH 0.032m APEX DIAMETER

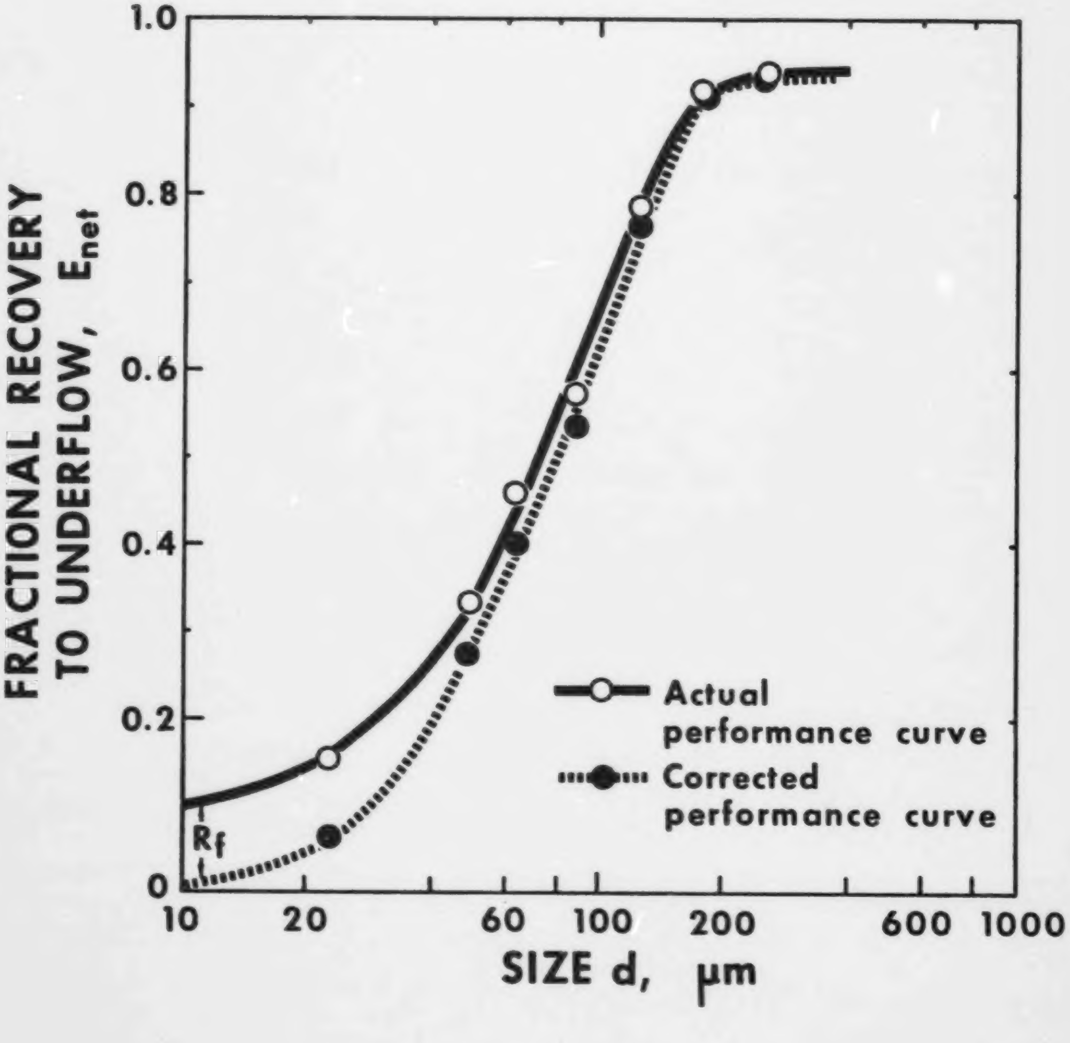


TABLE 3: COMPARISON OF ADDITIONAL SIMULATED AND ACTUAL RESULTS

	Actual	Simulated
$d_{50(C)}$ (μm)	80.0	47.1
U/F solids mass flow-rate (tonnes h^{-1})	146.5	186.0
U/F percent solids (%)	67.7	73.8
O/F solids mass flow-rate (tonnes h^{-1})	170.0	130.5
O/F percent solids (%)	18.9	14.5
Pressure drop (for given volume flow-rate) (kPa)	125	35

In order to provide a better description of the cyclone behaviour, the model constants were determined as a function of the most readily changeable design variable, apex diameter. To do this, cyclone performance was analyzed by evaluating R_f, m and $d_{50(c)}$. These were determined as follows:

(a) As shown in Figure 2, the recovery of fine particles to the underflow does not approach zero at small values of d. This is due (as described above) to short circuiting of fines to the underflow in direct proportion to the water flow in the underflow stream. R_f, therefore, is equal to the fraction of material unavailable for classification and is calculated as follows,

$$R_f = \frac{\text{water in U/F}}{\text{water in Feed}}$$

where water in U/F = $\dfrac{(100-\text{U/F percent solids})\ \text{wt\% recovery of solids in U/F}}{\text{U/F percent solids}}$

$$= \frac{(100-67.7)\ 46.4}{67.7} = 22.0 \text{ units}$$

for the cyclone operating with 0.032 m apex diameter. Similarly for the O/F;

$$\text{water in O/F} = \frac{(100-18.9)53.6}{18.9}$$

$$= 230 \text{ units}$$

Thus,

$$R_f = 0.087$$

(b) The actual performance curve (Figure 2) can therefore be amended to account for the by-passing of solids. The corrected recovery of solids for a given size fraction is given by

$$R_{\text{corrected}} = \frac{R_{\text{actual}} - R_f}{1 - R_f} \quad \ldots\ldots \quad 8$$

The corrected performance curve thus obtained is shown by the dotted line in Figure 2. As illustrated the curve tends to zero

FIGURE 3

REDUCED PERFORMANCE CURVE KREBS DIOLB CYCLONE WITH 0.032m APEX DIAMETER

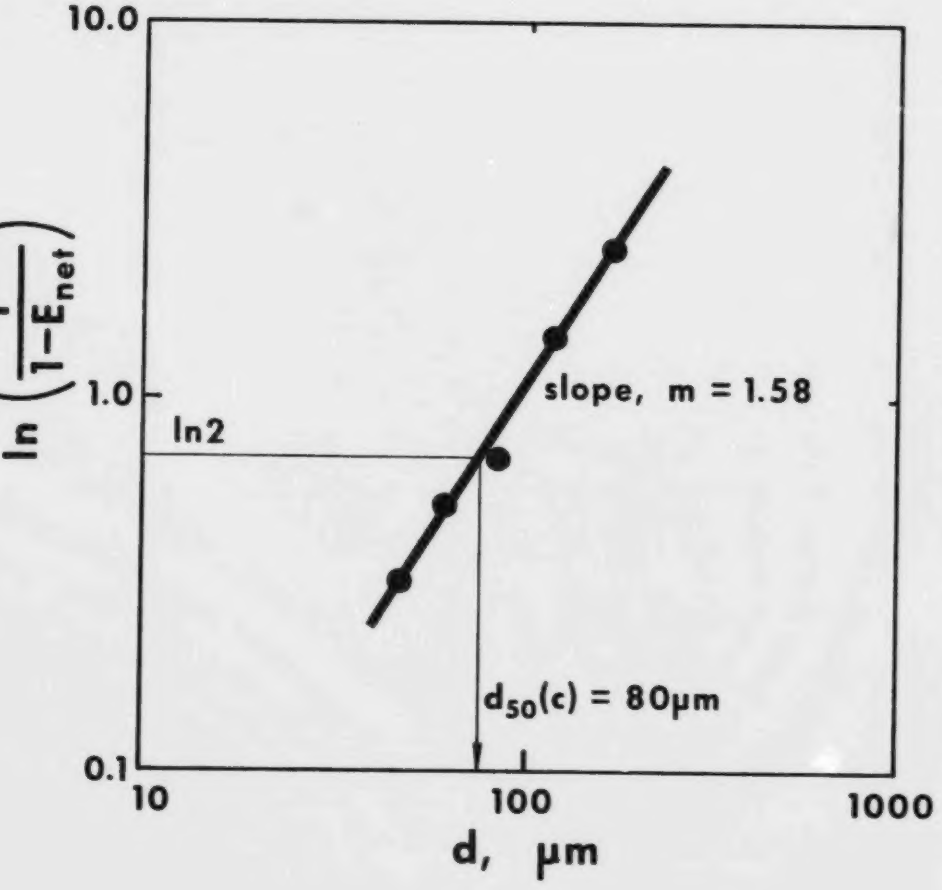
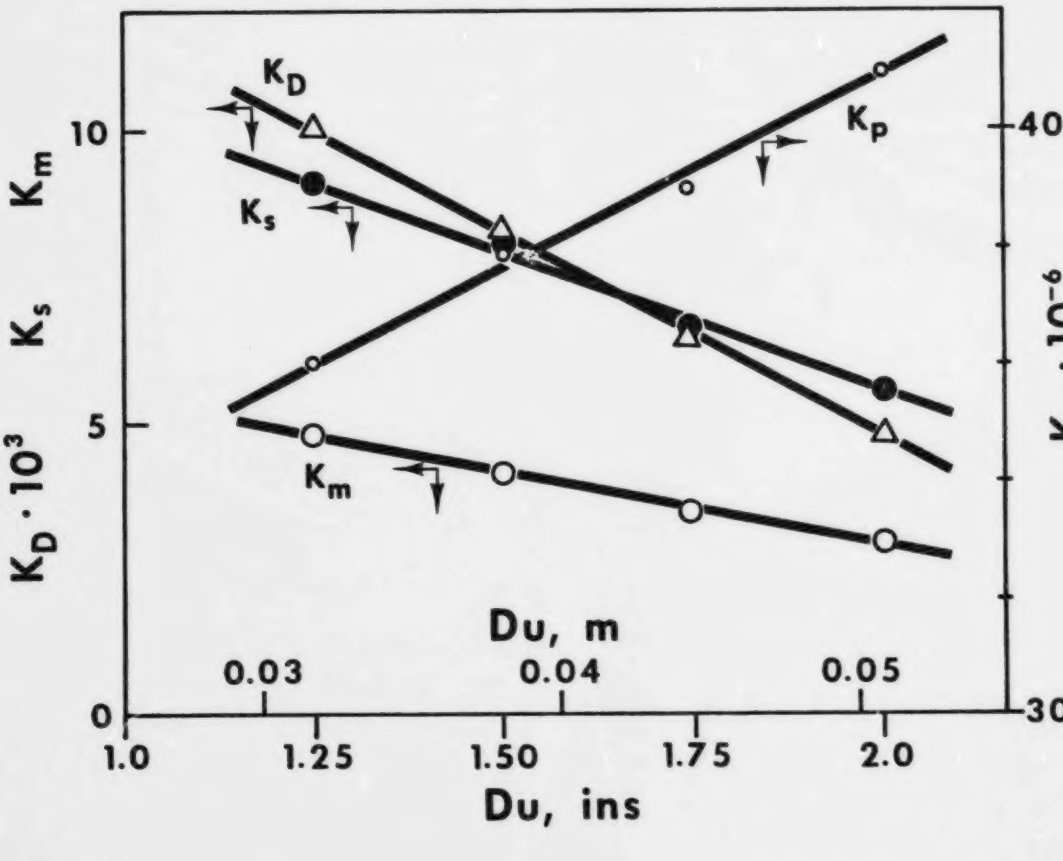


Figure 4

MODEL CONSTANTS AS A FUNCTION OF UNDERFLOW DIAMETER



at small values of d. The corrected cut size, $d_{50(c)}$, is by definition the size at which $R_{corrected} = 0.5$. It may be more reliably calculated from the reduced performance curve however.

(c) The reduced performance curve is given by equation 5.

$$E_{net} = 1 - \exp[-0.6931({}^{d/d}{}_{50(c)})^m]$$

where E_{net} is equivalent to $R_{corrected}$. This equation may be linearized to give

$$\ln \ln \left(\frac{1}{1-E_{net}}\right) = m \ln d + \ln \ln 2 - m \ln d_{50(c)} \quad \cdots\cdot \qquad 9$$

m is the slope of the straight line obtained by plotting ln ln $(1/1-E_{net})$ against ln d. Further, $d_{50(c)}$ occurs where $[1/1-E_{net}] = 2$. The reduced preformance curve for the conditions given above is shown in Figure 3 (in the curves as shown, $\ln(1/t-E_{net})$ is plotted against d, both on logarithmic axis).

The values of R_f, m and $d_{50(c)}$ for the apex diameters investigated are given in Table 4. These were then used to evaluate the equation constants, the resulting relationships are shown in Figure 4. In each case, a high degree of linearity exists between the constants and underflow diameter, as indicated by the correlation coefficients for the regression equations of the lines given in Table 5.

TABLE 4: 'B' SANDPLANT PRIMARY CYCLONES VALUES OF PERFORMANCE PARAMETERS

Apex Diameter (inches)	Flow-Ratio R_f	Sharpness of Separation m	$d_{50(c)}$ (μm)
2.0	0.21	1.00	28
1.75	0.18	1.13	39
1.5	0.12	1.37	59
1.25	0.08	1.58	80

TABLE 5: REGRESSION EQUATIONS FOR MODEL CONSTANTS (10" KREBS)

Equation	Correlation Coefficient
$K_p = 0.0003\ D_u + 0.00003$	0.998
$K_s = 14.87 - 183.86\ D_u$	0.981
$K_D = 18840.5 - 277299.2\ D_u$	0.996
$K_m = \exp\ (2.327-2.202\ D_u/D_o)$	0.997

It was also observed that the equation relating probability of a particle reporting to the cyclone underflow to particle size, equation 5, overestimated the recovery of the finest size fraction (-45 μm) in the underflow stream. This discrepency increased with increasing underflow diameter. As the fines fraction constitutes up to 40% of the B circuit sandplant feed, the differences shown are most significant. The error involved was corrected using the regression equation for the plant data,

$$E_{-45} = 315.8 \frac{D_u}{\%\ \text{solids}} - 0.2 \qquad \ldots. \qquad 10$$

for which the correlation coefficient was 0.988.

These modifications were incorporated into the model and the equations formed the basis of a hydrocyclone simulation program, SHYCYSIM. The effect of these changes on weight recovery to the primary underflow is shown in Table 6. Clearly an improved description of the cyclone classifying action has resulted.

TABLE 6: WEIGHT RECOVERY IN PRIMARY UNDERFLOW

Apex Diameter (inches)	Actual Recovery (%)	Predicted Recovery (%)	Predicted Recovery (Including Corrections)
2.0 (29.2% solids)	67.9	76.1	68.7
2.0 (34.8% solids)	60.7	76.0	63.5
1.75	63.6	66.9	62.6
1.5	54.4	47.0	55.2
1.25	46.1	46.0	45.9

CIRCUIT SIMULATION

Circuit simulation was carried out using the hydrocyclone program, SHYCYSIM, to predict the performance curves for each

Figure 5

MASS BALANCE FOR NOVEMBER 1977 'B' SANDPLANT CIRCUIT

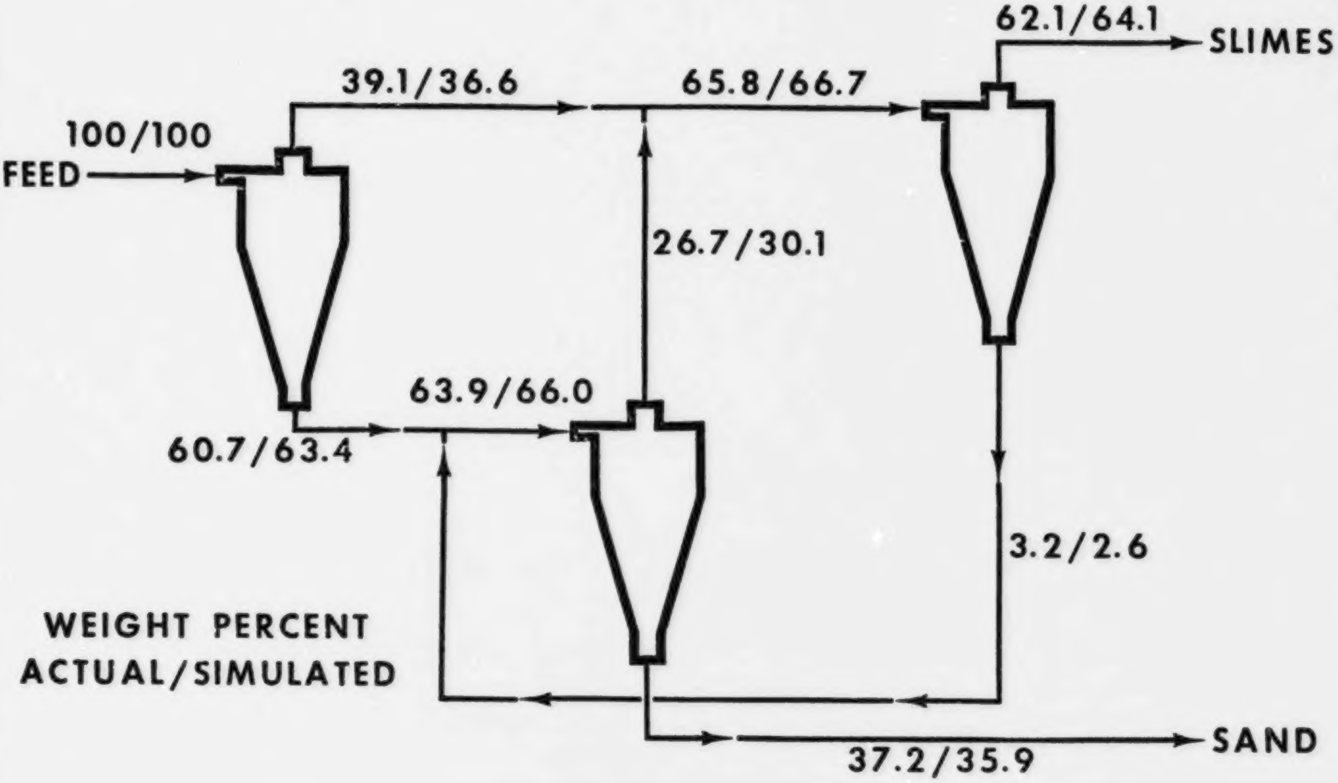
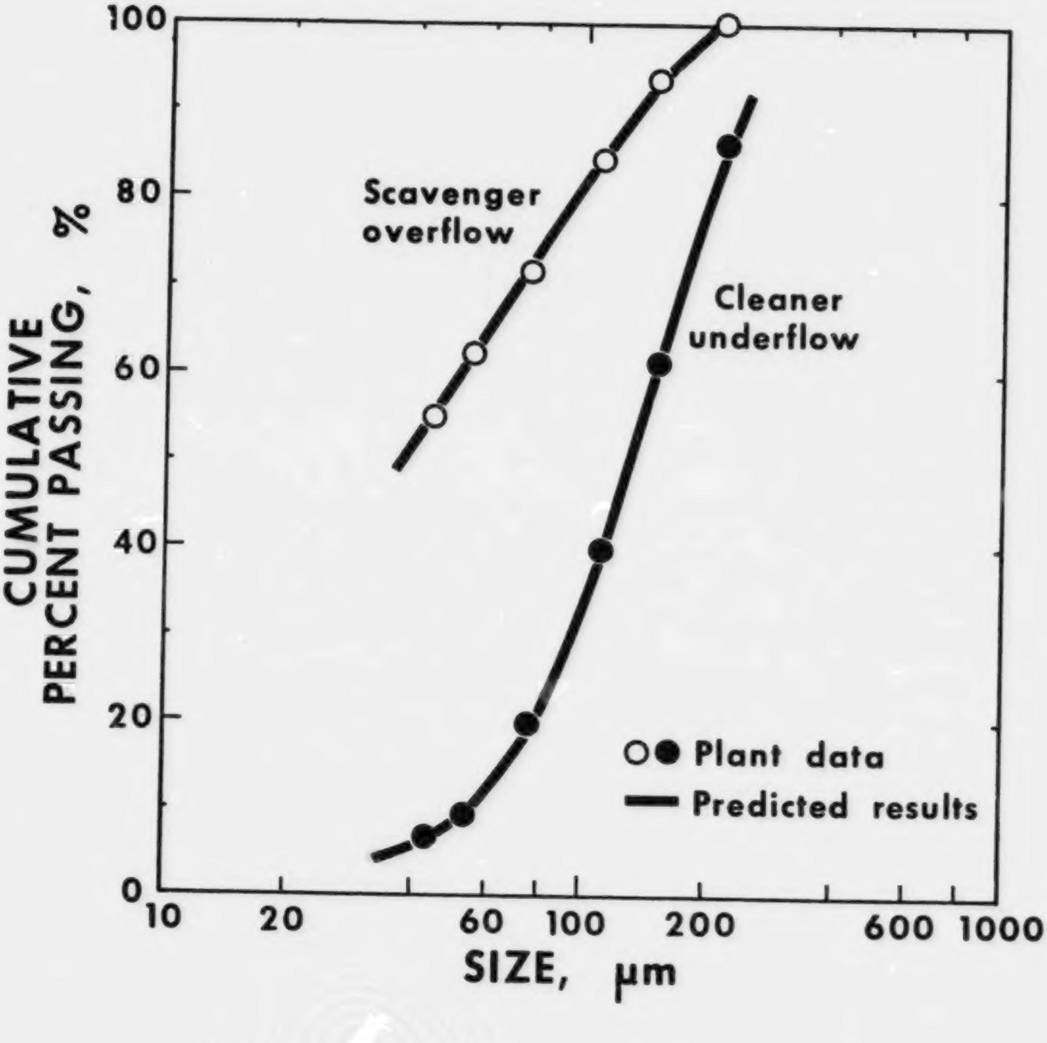

62.1/64.1
SLIMES
39.1/36.6
65.8/66.7
100/100
FEED
26.7/30.1
63.9/66.0
60.7/63.4
3.2/2.6
WEIGHT PERCENT
ACTUAL/SIMULATED
SAND
37.2/35.9

Figure 6

PRODUCT SIZE DISTRIBUTIONS FROM NOVEMBER 1977 'B' CIRCUIT OPERATION



cyclone bank under any set of operating conditions. These results were then used as input data to a modular computer program, MPFLOT, which uses an iterative technique to arrive at a steady state balance for the circuit. The program, MPFLOT, has been described previously in connection with predicting locked cycle flotation test results from batch data (6).

The 'B' circuit sandplant was simulated under the conditions operating in November, 1977, as shown in Table 7. The material balance for the circuit is shown in Figure 5. There is excellent agreement between predicted and actual mass flows in all streams. The predicted product size analyses are compared to plant results in Figure 6. Again, there is good agreement between the two sets of data.

TABLE 7: NOVEMBER 1977 CIRCUIT

Design Parameters

Bank	Type	Number	Cyclone Diameter (m)	Vortex Finder Diameter (m)	Free Vortex Height (m)	Inlet Area (m²)	Apex Diameter (m)	Cone Angle (deg)
Primaries	Krebs	8	0.25	0.089	0.79	0.020	0.045	22
Cleaners	INCO	8	0.31	0.150	1.91	0.031	0.076	18
Scavengers	"	8	0.31	0.150	1.91	0.031	0.076	18

Operating Parameters

	Pressure drop (kPa)	% Solids	% -45 μm (feed)
Primaries	150	30	35.7
Cleaners	100	45	14.5
Scavengers	150	25	47.5

FIGURE 7

EFFECT OF FEED PERCENT SOLIDS ON WEIGHT RECOVERY TO SANDFILL

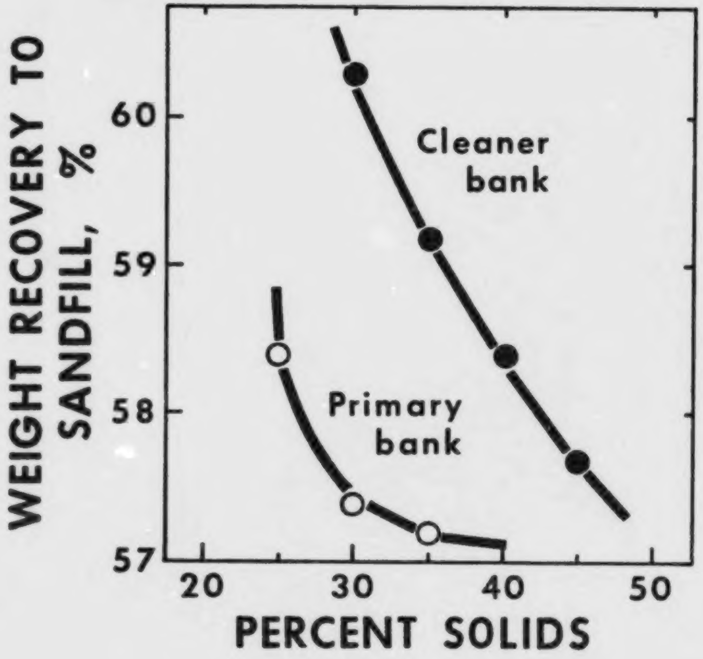
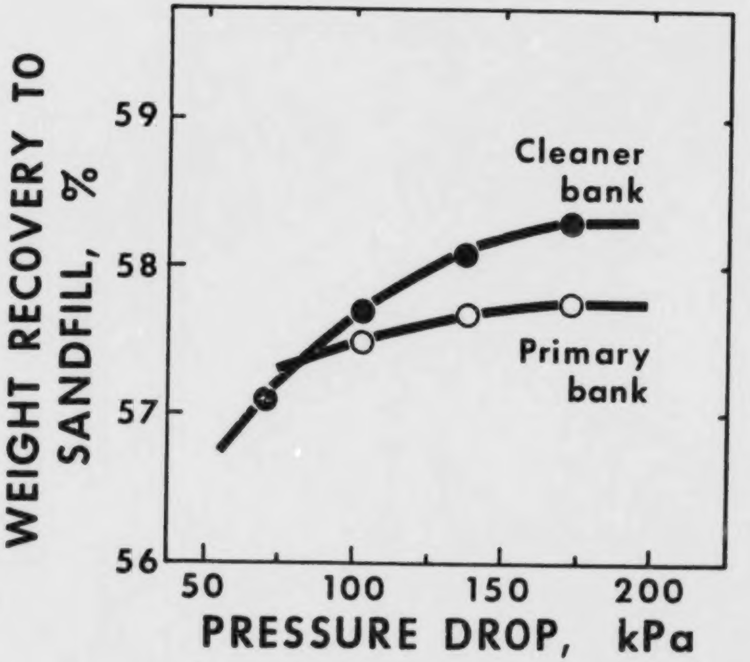


FIGURE 8

EFFECT OF PRESSURE DROP ON WEIGHT RECOVERY TO SANDFILL



CIRCUIT OPTIMIZATION

As previously described, order to increase throughput and provide more efficient recovery of coarse particles to the primary and cleaner underflows, 10" Krebs cyclones were installed in the cleaner banks and additional 15" Krebs cyclones added to the primary bank. The best of the cyclones that were replaced were then installed in the scavenger bank. With the cyclone design variables fixed, the effect of changes in the principal operating conditions were investigated in order to determine the conditions required for optimum circuit performance.

The effect of percent solids in the primary and cleaner feed streams on weight recovery of solids to sandfill is shown in Figure 7. It is apparent that feed percent solids to the primaries has only a minor effect on recovery to the final product. This variable is more significant in the case of the cleaners, where as percent solids increases, solids recovery to the cleaner underflow decreases. This decrease, however, is largely due to increased recovery of -45 μm material in the overflow stream and is therefore not entirely disadvantageous.

The effect of pressure drop across the primaries and cleaners is shown in Figure 8. As the pressure drop across both cyclone banks increases, weight recovery to cleaner underflow also increases. The change is, however, slight for both banks.

Underflow orifice diameter has the most significant effect on cyclone performance. Weight recovery to sandfill is significantly reduced when a small apex diameter is used. The effect is clearly illustrated in Table 8.

TABLE 8: EFFECT OF CLEANER APEX DIAMETER ON WEIGHT RECOVERY TO SANDFILL

Cleaner Apex Diameter (m)	Weight Recovery To Sandfill
0.051	57.7
0.045	53.6

For optimum circuit performance, therefore, it would appear that the most important operating factor is to maintain as high a water dilution in the feed streams to each bank within the constraints of sump and cyclone capacity, and the quantity of fines acceptable in the sand fill product. For the B circuit the optimum conditions within the constraints mentioned were; 30% solids to the primary bank, 30-40% solids to the cleaners (depending on the throughput) and 20% to the scavengers. Pressure drop was arbitrarily maintained at 140 kPa. The apex diameter recommended for the cleaners was

0.045 m, the maximum size permissable without excessive dilution of the sand fill with fines.

The ease with which the circuit simulation program can be manipulated means that various circuit configurations can be readily simulated. For example, in order to increase the coarseness of the sand fill, operators often direct the primary and cleaner overflows directly to tails without scavenging. This change was made in the modular program and the method of operation simulated. The results indicated that there was a slight decrease in the percent -45 μm fraction in the sand fill, from 19.3% to 14.7%, but this was accompanied by a drop in weight recovery from 50.4% to 43.0%. This prediction was confirmed in the plant, and therefore this mode of operation was considered best avoided.

Similarly, the effect of a regrind stage upstream of the sandplant can be determined by altering the feed size distribution to the circuit. For a change in fineness of grind from 32.1% to 45.4% -45 μm, as was recently implemented, a decrease in weight recovery of 12.8% would be expected if the operating conditions remained unchanged. The actual decrease was observed to be 13%.

Since the above work, new Krebs cyclones have been installed in the sandplant circuit treating Pipe Open Pit tailings

(C circuit). This material is ground much finer than Thompson Mine Ore, and the tailings slurry is extremely viscous due to the presence of substantial quantities of asbestos. The model developed above was applied to C circuit sandplant, and good agreement found between predicted and plant results despite the differences in the feed material. The feed percent solids to all banks was maintained at 15% or less (to minimize viscosity effects), the remaining conditions being similar to those in the B circuit sandplant. Under these conditions it was predicted that weight recovery to sandfill would be 50.4% with 19.3% -45 µm. Plant weight recovery was found to be approximately 45% with 15% -45 µm (7).

CONCLUSIONS

The model of the hydrocyclone presented here has been shown to describe the behaviour of the B Circuit sandplant cyclones accurately. The program, thus developed has been used in conjunction with a modular computer program that simulates various circuit configurations to predict the effect of changes in a number of operating parameters on circuit performance. It has been shown that weight recovery to sand fill increases as feed percent solids decreases, and, to a lesser extent, as pressure drop across the banks increases. Cleaner cyclone apex diameter has the most significant effect on weight recovery causing a rapid decrease in recovery as diameter decreases. The predicted changes in circuit performance are in good agreement with those observed in the plant. Similarly, changes in circuit configuration and upstream modifications in the size distribution of the ore can be quantitatively assessed.

The model has also been shown to be applicable to widely different types of ore.

The superiority of the model over a simple partition curve description of cyclone behaviour suggests that it is suitable for incorporation into grinding-classification simulation programs.

ACKNOWLEDGEMENTS

Permission from INCO Metals Co to publish this work is gratefully acknowledged. We should also like to recognize the contribution of R. Tipman and C. Edwards who were responsible for carrying out the sampling campaigns, D. Burgess who typed the paper and N. Hansen who drafted the figures.

REFERENCES

(1) Devlin, J.B., and Goddard, C.N., Sandfill Preparation and Slimes Disposal at INCO's Thompson Mine, CIM Reporter, 4(6), 1978, 12-16.

(2) Lynch, A.J., and Rao, T.C., Modelling and Scale-up of Hydrocyclone Classifiers, 11th I.M.P.C., Cagliari, 1975, 246-269.

(3) Plitt, L.R., A Mathematical Model of the Hydrocyclone Classifier, C.I.M. Bulletin, December, 1976, 114-123.

(4) Plitt, L.R., The Analysis of Solid-Solid Separations in Classifiers, C.I.M. Bulletin, April, 1971, 42-47.

(5) Schultz, N.F., Separation Efficiency, Trans SME/AIME, 247, 1970, 81.

(6) Agar, G.E., and Kipkie, W.B., Predicting Locked Cycle Flotation Test Results from Batch Data, CMP Annual Meeting, Ottawa, January, 1976, 357-393.

(7) Edwards, C.R., Thompson Mill Process Technology, private communication, May, 1978.

NOMENCLATURE

$d_{50(c)}$ = corrected cut size (μm)

D_c = diameter of cyclone (m)

D_i = diameter of (m)

D_o = diameter of vortex finder (m)

D_u = diameter of apex (m)

E = probability of a particle in feed reporting to underflow

h = free vortex height (m)

$$\left.\begin{array}{l} K_D = \\ K_M = \\ K_p = \\ K_s = \end{array}\right\} \text{constants}$$

m = sharpness of separation

P = pressure drop across cyclone (kPa)

Q = volumetric flow rate of feed slurry (m^3h^{-1})

R_f = flow ratio

S = volume split

η = separation efficiency

ρ = density of liquid (tonnes m^{-3})

ρ_s = density of solids (tonnes m^{-3})

Ø = volume fraction of solids in feed slurry

Discussion of Paper by R. Stratton-Crawley and G. Agar

R. Faucher: Did you try to adjust the water ?

Reply: No, the votex height was kept constant.

L. Plitt: Did any of the test conditions approach rope discharge in the underflow ? How valid are your equations as they approach these conditions ?

Reply: This is probably the reason for deviation from your original equations. Your equations were developed for optimum conditions under free discharge, whereas in the plant to obtain a high pulp density of discharge, the operators tend to operate towards rope discharge.

A. Mular: How does you computer simulated program handle the real cyclone entering the real roping condition ?

Reply: I think that it is largely taken into account by my correction for the over estimation of fines by the reduced efficiency equation and the effect that roping could have been to selectively eliminate fines from the underflow, whereas under roping conditions they may have been crowded out. That is why the correction is needed. The program doesn't know if the cyclone is roping or not.

K. Konigsmann: Do you think that your program would permit you to predict the performance of other tailings ?

Reply: Without the regrind the product contains 40% minus 45 microns.

K. Konigsmann: Does your program take into account pulp streams higher in sulphur or containing much finer material ?

Reply: In a test with a stream of asbestoes where the pulp viscosity was much higher than the stream on which we developed the equations, there was quatitative agreement between actual and predicted results. We have used it on a limonite ore which was exceedingly finer and again got good agreement. It depends mainly how the cyclones are operated.

J. Palmer: If you have mixed particle sizes in the feed, is there any problem in predicting the results ?

Reply: These equations were developed for a complete particle size distribution.

Paper No. 6

THE CANADIAN BULK CARGOES COMMITTEE

by

Klaus Konigsman

INTRODUCTION

Representatives of various industries and the Canadian Coast Guard formed a Bulk Cargoes Committee three years ago. In 1976 and 1978, I have attended the meeting which take place once per year here in Ottawa as representative of the Canadian Mineral Processors, Dave Hendricks from Granisle accepted the charge in 1977. Our Secretary, Bob Bruce, has suggested repeatedly that I report to the Membership on the deliberations fo this group. Twice I found good excuses; this year I did not. I was reluctant to address you since the Committee's deliberation proceeded at a pace which might be identified as Government squared. The Committee advises a Government agency which in turn negotiates with foreign Government agencies.

At the outset, I will describe the developments which led to the formation of the Committee and its purpose.

The Canadian Mining Industry ships annually millions of tons of products to overseas customers; be it iron ore, coal, fertilizer, or base metal concentrates. Most loadings are in vessels of foreign registry and all will be directly affected by Canadian or International regulations of the transport of such bulk commodities. The major shipping nations are members of an Intergovernmental Maritime Consultative Organization (IMCO). This organization elaborates regulations which govern shipment of all cargo, bulk materials and concentrates included. At IMCO'S deliberations, Canada has one vote, one voice on the Consultative Committee and is represented by our Coast Guard. The Coast Guard has over the years been in contact with industries when discussions concerned shipment of bulk cargoes. In the early sixties, a "Concentrate Committee" did exist. It consisted of metallurgists from a number of mining companies and Government representatives. This committee was instrumental in the development of what became known as the "Concentrate Code", legislation which has evolved into the Canadian Bulk Cargoes Code as it is known today. The concepts of the flow moisture point and transportable moisture limit originated from its work. Incidentally, both criteria are widely accepted and utilized world wide. The Concentrate Committee faded away once the Concentrate Code has been adopted.

Shipping, like any other enterprise, is constantly changing. Ship sizes increase. Tonnages handled have risen dramatically, one needs to think only of iron ore and coal developments in the last decade. New products are to carried by vessels, the most striking example is ferro silicone. This material gives off large volumes of hydrogen when it comes in contact with water and is therefore a rather explosive cargo to carry. Consideration for safety of crew and ships have been emphasized after some mishaps. In consequence of such developments, regulations affecting bulk cargo are continually evolving, and the shipping of bulk commodities has not yet been under attack from environmentalists.

By the mid-seventies, evolution of shipping regulations proceeded in a direction which could have led to drastic cost increases for export of mining products. One example was the classification of concentrates as dangerous cargo. The industry was poorly informed about and not involved in the elaboration of rules. A Bulk Cargoes Committee of the Marine Safety Advisory Council was formed in 1976 in the attempt to re-establish government and industry collaboration. Chaired by the Coast Guard it draws members from the iron ore companies, the Canadian Coal Association, shipping companies, the Research Council and the Canadian Mineral Processors. As mentioned, I have been a representative of our association. This much on the history and purpose of the Bulk Cargoes Committee.

In the two years during which I attended its sessions, the discussions were primarily about changes in regulations which affect sulphide concentrates and I will restrict my report to this topic. Shipment of iron ore, coal and other bulk commodities were also discussed.

Three topics have been given considerable attention:

1. The transportable moisture limit and the determination of the flow moisture point.
2. Sampling procedures for dockside stockpiles is a matter which concerns shippers of iron ore even more than those of base metals.
3. Oxidation and self heating of concentrates.

The first two topics have been around for a long time, and yet changes of regulations continue to be proposed. The initiative may come from some national interest once investigators of other nations have developed improved test methods. Frequently they are the result of recommendations made by boards of inquiry into shipping mishaps. Two types of accidents have led to proposals for change. The shifting of a ship's load of wet concentrate or coal makes it list severely and it may be compelled to head for a part of refuge. Several smaller freighters capsized in European waters from such cause. The reliablity of flowmoisture determination and sampling methods are questioned. Oxidation and self heating has been enocuntered by Australian investigators with concentrates from their continent. There was also one mishap caused by a Canadain shipment.

A detailed listing of the proposals made by various countries would be long and tedious. The discussions may be summarized under the following headings:

(1) Transportable Moisture Limit

Flow moisture and transportable moisture limit are currently defined by procedures which were developed in Canada years ago. You are most likely familiar with the determinations. A concentrate sample is compressed into a mold having the form of a truncated cone (4" dia. - 2" high). This cone-shaped sample is placed on a test table which drops 1/2 inch at a frequency of 25 stroked per minute. If the sample is dry, it crumbles. It has reached the flow moisture once plastic deformation of the sample is apparent. The transportable moisture limit thus defined has been proven safe and is probably quite conservative. The bulk cargoes comittee has raised objections to different test methods proposed by Norway which, in effect, would have lowered the transportable moisture limit. This limit is of critical importance to all Canadian shippers, particularly those who do not dry their concentrates. Even a drop of one percentage point in the TML would be a serious matter. The Canadian position has been to defend the flow moisture determination. It is reliable and widely recognized as such. This argument has apparently been accepted. At the Committee meeting in 1978, one gained the impression that the matter had finally been laid to rest.

(2) Sampling of Stockpiles

The debate on this topic has at times been lively. It is impossible to sample stockpiles of 40 to 80,000 and over 100,000 tons by applying the same neat rules which govern sampling of small quantities. Rarely can one get onto a pile for fear of being drowned in a slide. One cannot drill into it or pipe sample such a pile and yet rule-makers ant a " representative sample " taken prior to leading. Arguments by the industry have been that a stockpile's composition and moisture does not change significantly, that oxidations and precipitation affect only a thin layer at the pile's surface. This view has at least been accepted by the Canadian Coast Guard.

It is recognized by them that sampling is adequate if carried out in railroad cars, trucks or on conveyors prior to stock piling. Sampling has to be repeated for moisture on the conveyor loading a ship. For the quick moisture determinations during loading readings of the Speedy Moisture Tester have been approved. I do not believe that this solution to the sampling controversy has made it over all International hurdles. But then, IMCO and the Bulk Cargoes Committee have to be kept occupied.

(3) Oxidation and Self Heating of Concentrates

On the initiative of an Australian delegation to IMCO conferences, all sulphide concentrates were classified as " dangerous cargo " in 1975. This classification resulted from an Australian study which had concluded that certain sulphide concentrates would absorb the oxygen in the ship holds and constitute a risk for crew and ship. It was claimed that there was a risk of spontaneous combustion. The Australian study was based on a few isolated incidents. The temporary classification as dangerous cargo caused additional costs in a few shipments from the Canadian East Coast where port wardens insisted on having the concentrates levelled and covered with plastic sheeting to reduce oxygen flow to the load. The classification was modified in 1977. Had it been maintained, the additional costs for insurance and unnecessary safety measures could potentially have been considerable to all shippers. The dangerous cargo classification of concentrates has been replaced by a new one, namely " material hazardous only in bulk ". Even this wording appears extreme where it refers to concentrates, but it was the least offensive which Canadian representatives could get accepted. Under the listing " hazardous in bulk " and in the concentrate sub-entry, one does still find reference to the characteristics which gave rise to the dangerous cargo entry (oxygen absorbtion and self heating) but at the end of the section there is one paragraph which permits exempting from all restrictions concentrates which are known to be harmless. Almost all Canadian concentrates can thus be exempted.

This summarizes the achievements of the Bulk Cargoes Committee in the past three years.

- Definition of the transportable moisture limit remains unchanged
- Reasonable sampling procedures are accepted by the Coast Guard
- Concentrates are not listed as "dangerous cargo" but are in the less severe classification of "hazardous only in bulk" from which most Canadian products are exempt.

One cannot claim that the results are revolutionary and it might be asked whether or not the effort is worthwhile. I would think it is. Regulations are likely to be changed in the future and the mining industry should participate in their evolution.

Discussion of Paper by K. Konigsmann

D. Bridges:	Is the Bulk Cargo Committee concerned with the shippers per se, regulation and safety, or what ?
Reply:	The B.C.C. advises the coastguard only. The transport of concentrate is regulated by the Bulk Cargo Code and is constantly revised because of accidents which occur. The committee has no legislative power.
B. Neal:	Have there been any standard tests done to define hazardous material due to heating ?
Reply:	No, there are no accepted standards but there are two tests. Germany has carried out a great number of tests on sulphide concentrates. Quantities of about 5 kilograms are heated to 300 degrees C and they noted that the temperature kept rising. The results have been taken into account but is not accepted as a standard test. Noranda has been working on tests where the temperature rise is recorded when oxygen is supplied to the sulphide concentrate. This is also not accepted as a standard test.
D. Pickett:	How do you test for moisture of a mixed cargo iron ore ?
Reply:	Soil stability tests have been applied to iron ore successfully.

Paper No. 7

OPERATION & MAINTENANCE OF GRINDING CIRCUTIS AT SIDBEC-NORMINES

by

Michael Jerome Wallace*

INTRODUCTION

This paper refers to the wet open circuit ball mills in operation at Sidbec-Normines's pelletizing plant at Port-Cartier, Quebec, which commenced operations in late November 1977.

Prior to detailing our experiences in the grinding area at Port-Cartier, a general description of the total Sidbec-Normines project is outlined.

HISTORY

Sidbec-Normines was incorporated in 1976 to develop the Fire-Lake - Port-Cartier iron ore mining, concentrating and pelletizing project. The primary objective of Sidbec-Normines is to produce 6 million tons per year of oxide pellets, of which 3 million tons per year are of a low silica grade suitable for direct reduction purposes. The participants in the project are Sidbec, the Quebec Government steel producer, British Steel Corporation (Int'l) and the Quebec Cartier Mining Company. Equity interest is held in the company by the participants in the proportions of 50.1%, 41.67% and 8.23%, respectively.

The project included the purchase of the Fire Lake orebody, the town of Gagnon and the existing Lac Jeannine concentrator.

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* Maintenance Superintendent, Pelletizing Plant, Port Cartier, P.Q.

FIRE LAKE / LAC JEANNINE

The Fire Lake mine is located 100 km (62 m) northeast of the town of Gagnon, Quebec, and the crude ore is railed from here to the old Lac Jeannine concentrator. The concentrate produced there is then moved by rail 300 km (186 m) south to the Port-Cartier pelletizing facility.

The development and operation of the Fire Lake open-pit mine and the continued operation of the Lac Jeannine concentrator are being carried out on behalf of Sidbec-Normines under a management contract agreement with Q.C.M.

The Lac Jeannine concentrator and its grinding circuits have been the subject of many technical papers during their 17-year life and consequently will not be detailed here.

PORT-CARTIER

From Lac Jeannine, concentrate is railed to the S.N.I.'s pelletizing complex located immediately east of the town of Port-Cartier.

Construction of the pelletizing complex began in May 1975 and initial pellet production commenced on Line 1 in late November 1977. The complex consists of a secondary concentrator and pelletizing plant containing facilities for grinding, filtering, additive mixing, balling and induration of pellets. Processing equipment is conventional except for the secondary concentrator which uses wet high intensity magnetic separation in the first large scale application of the Jones units in the North American iron ore industry.

MATERIAL FLOW

In Line no. 1, dry feed from three bins is either fed directly to three ball mills or on demand is diverted, by reversing the belts, to feed the secondary concentrator, where the silica content of 3,000,000 tons of concentrate annually can be reduced to less than 2.5% SiO_2 by wet high intensity magnetic separation from a nominal 5.8% SiO_2.

In the secondary concentrator, the material is fed with dilution water into one of six Jones wet high intensity magnetic separators rated at 80 tons of new feed per hour. From the Jones machines, the tailings are pumped to a tailings pond, and the product is recovered at the rate of 413 tons per hour of solids for all six units and pumped to the ball mills on line 1.

In Line no. 2, standard silica concentrate is fed from the other three storage bins directly to the three ball mills. The grinding section incorporates six 5.1 m (17 ft) diameter by 10.3 m (34 ft) long wet ball mills, each powered by a 6400 HP single pinion drive low-speed synchronous motor and operating in open circuit.

The discharge from the grinding circuit is pumped to slurry storage tanks and from there the slurry is distributed to the filtering section for dewatering in Eimco vacuum filters arranged in banks of five units on each line. The filter cake, containing a nominal 8.5% moisture, is conveyed to a central filter cake bin in each of the two lines. After withdrawal from the bins, the filter cake is mixed with bentonite and dolomite in a Littleford mixer in each line.

BALLING & INDURATING

The mixture is fed to four 7.5 m (24.6 ft) diameter Dravo balling discs in each line. From the discs, the green balls having a nominal 13 mm (1/2 in) diameter are fed by a Sala reciprocating conveyor at the rate of 420 tons per hour in each line to standard 4 m (13 f) wide by 116 m (380 ft) long Dravo-Lurgi indurating furnaces with 58 windboxes. These machines have the same dimensions as those installed in 1976 at Hibbing Taconite Co. and in 1977 at the Minorca Mine of Inland Steel Mining Co., both in Minnesota.

The green balls pass through the drying and prefiring zones prior to being fired at 1340°C (2450°F). The fired pellets are cooled and discharged into pellet surge hoppers, from which they are conveyed onto stockpile via a Stephens Adamson 2000 tons/hr stacker.

Shiploading is accomplished via a Stephens Adamson bucket wheel reclaimer rated at 7,200 tons/hr.

That concludes a general description of the Sidbec-Normines project.

1978 PRODUCTION

The 6 ball mills were commissioned between November 1977 and August 1978 in sequence from "A" to "F".

Consequently, mill "A" has produced the most ground ore to date with a total tonnage at year end 1978 of approximately 430,000 tons and mill "F" being the last one to be commissioned has produced only approximately 120,000 tons of ground ore.

During 1978, pellet production for the 9 months of operation was 1.9 million tons. Note also that production was suspended for 3 months in 1978 because of our inability to obtain concentrate from Q.C.M. during the strike of the United Steelworkers Union. Only one of the induration machines was operated during the year but all 6 mill discharges were interconnected in a common header. This enabled us to commission all the mills, and provided the advantage of having 6 mills available for one line so that production was not reduced as a result of poor mill availability.

Nominal monthly design production of 250,000 tons/month was achieved in September 1978 and exceeded by 22% in October 1978 when 307,000 tons were produced.

During the year, a total of 600,000 tons of low silica pellets were produced. Mills "A", "B" and "C" were originally supplied with cyclones to handle the slurry pumped to the mills from the secondary concentrator. During the year, a cyclone was added to mill "D" so that low silica pellet production would not be reduced if mills "A", "B" or "C" were unavailable.

GRINDING

Ball mill feed, as mentioned above, is either direct from the bin storage with a nominal 2% moisture or in the case of low silica feed is in the form of a slurry of 60% density via a cyclone.

Concentrate feed has the following specifications: 90% -20 mesh, 65.0% Fe and 5.8% SiO_2. One composite sample is taken of each train from Lac Jeannine for chemical, structure and moisture analysis.

Feed rate of approximately 140 tons/hr and process water addition are adjusted to give a discharge with the following specs: 65% -325 mesh, 1600 blaine and 70% density. Discharge slurry is sampled every 4 hours and analysed to check the percentage -325 mesh. Results are phoned to the Control Room within one hour and tonnage adjusted as necessary.

A correlation has been achieved between blaine and -325 mesh and now blaine analysis is only done on request. Density is checked by the operator every hour who phones the results to the Control Room.

MILL COMPONENTS

Taking the ball mills component by component, our experiences have been as follows:

A - Drive Motors & Controls

The 5.1 m (17') x 10.3m (34') long mills are driven by General Electric 6400 HP synchronous motors running at 164 RPM (163.6 exact) with a motor service factor of 1.

No major problems were encountered during the start-up of the motors except where some exciter windings had been damaged during handling by construction personnel. These windings were repaired on site under supervision of G.E. personnel.

The motor cooling via an air to water heat exchanger has presented some problems because of the fact that our cooling water - which is, in fact, recycled process water - has on occasion run at temperatures above the design maximum of 27°C (80°F). RTD's have been relocated and trip points on air temperatures reset with G.E.'s approval in order to enable us to cope with these slightly higher temperatures.

Controls consist of a local alarm panel located on the ground floor under each mill, and all temperatures can be scanned on a temperature multipoint indicator in the MCC Room. Kilowatt meters and scale readings are located on the control room panel.

B - Clutch

The motor drives the pinion via a Fawick airflex dual 60 VC 1600 clutch. Air supply is at 8000 KPA (115 PSI) and time to engage the clutch is 3-4 seconds. On engagement, air pressure drops by 125 KPA (18 PSI) with recovery to the original pressure in 3-4 seconds.

Recent discussions with Fawick indicate that the air receivers originally supplied to us are too small to permit one safe start of the mill if the air supply has failed. They have supplied us with an air receiver of triple the capacity of the existing unit. This should also reduce our pressure drop during mill starting.

Engagement and disengagement of the clutch is via a Wabco 3-way solenoid valve mounted on top of the air receiver. The air receiver and solenoid units have all been relocated from the ground floor under the mill motor to the operating floor beside the motor. This provides a cleaner environment for these units.

Sticking of solenoid valves became a problem after start-up, possibly because of dirt in the air lines and/or the frequent use of plant air for this function instead of the cleaner instrument air, as invariably occurs during start-up. This will also occur, of course, when the instrument air compressor is on service or being repaired. Solenoid valves have been changed to a slightly different model of Wabco valve with a housing that will reduce the frequency of sticking. The solenoids are also now lubricated regularly with 'lubriplate' grease. We are still not satisfied with this valve and while searching for a better alternative, we have incorporated an emergency automatic mill motor shutdown, which will activate and stop the motor in the event that the clutch has failed to disengage 10 seconds after receiving a signal to do so.

C - Drive Pinion & Ring Gear

The single drive pinion is of a helical configuration with 24 teeth and a face width of 1 metre (39 inches). This drives the gear at 13.9 RPM and is one of the biggest pinion drives of this nature in the world.

The ring gear is bolted together from 4 equal 90° segments and in turn is bolted to the mill head, its flange also being accurately machined to fit a spigot on the drive head.

Lubrication of the gear and pinion is via a Trabon open gear system, which operated initially with Whitmore and is now running on Keystone open gear.

Considerable attention was paid to accurate alignment of the pinion and gear during construction and to date no problems have been experienced in this area.

The pinion and ring gear are symmetrical in order to permit reversal for extended life.

D - Mill Liners & Discharge

All mill liners are rubber with Nihard lifter bars with the exception of mill "F" which has an all rubber discharge head.

Mills "B", "D" & "F" are fitted with Skega rubber, whereas "A", "C" and "E" are Trelleborg.

All liners are bolted through the shell with 32 mm (1¼ in) diameter square head bolts incorporating a steel cup washer with a 6 mm (¼ in) rubber insert. Shell liners are 60 mm (2-3/8 in) thick rubber plates with 90 mm (3½ in) Nihard lifter bars; head liners are 50 mm (2 in) thick rubber plates with 75 mm (3 in) Nihard lifter bars.

Wear on mill liners and lifter bars was measured on all mills at year end 1978. Naturally mill "A", having produced the most ground ore, also exhibits the most wear.

As mentioned, the mill feed and discharge head liners are of radial rib design, Nihard casting lifter bars with 50 mm (2 in) thick rubber plates.

The lifter bars on the head of mill "A" close to the feed chute display the most wear. The wear in this area is approximately 10% of lifter bar thickness per 100,000 tons. This mill will be watched closely as its tonnage reaches the ½ million mark and the lifter bars reach approximately 50% worn. It is anticipated that these lifter bars on the feed end head will be changed out at between 5 to 600,000 tons of through-put.

Mill heads are of cast Meehanite with the trunnions cast integral with the heads. The discharge heads are fitted with 2 m (6 ft) long bolted steel construction trommel screens with slotted openings 9 mm (3/8 in) x 25 mm (1 in) long. The wearing surface and retarding spiral are lined with rubber.

The discharge is pumped from a small discharge box under the trommel screen to slurry storage via an ASH 6 x 6 x 17 pump with Nelson variable speed fluid drive.

Unexpected mill stoppages result in this box overflowing which creates a clean-up problem under the mills. We have increased the capacity of the sump pumps under the mills to handle this. If the operator stops the mill, he can shut off the feed a few minutes beforehand, in which case the box does not overflow.

As we have reached steady operation, the number of unanticipated stops has decreased; however, we intend to change out these discharge boxes to larger capacity units in order to reduce the clean-up labour required in this area.

E - Grinding Media

Grinding is accomplished by means of 38 mm (1½ in) diameter forged steel balls with mills "C" & "F" operating on 38 mm (1½ in) slugs supplied by Quebec Iron Foundries, of Mont Joli, Quebec.

Most of our grinding balls are produced locally in Sept-Iles by Norforge, using 28 mm (1-1/8 in) x 9 m (30 in) rods supplied by Sidbec-Dosco.

This arrangement of production of grinding balls from Sidbec Steel was concluded during the construction of the pellet plant in Port-Cartier. Norforge, a subsidiary of Niagara Steel of St. Catherines, Ontario, then constructed a forging plant in Sept-Iles with a capacity of 60 tons/day of balls.

Delivery from Sept-Iles commenced in August 1978. Prior to that, balls had been purchased from Burlington Steel. Occasional shipments are still purchased from Burlington in order not to be dependent on one source of supply for balls.

The original system of a small loading bin of 2½ tons capacity, loaded from a storage bin which in turn was loaded from a pit by a magnet on the overhead crane, has been virtually eliminated. Twenty larger bins of 10 tons net capacity were built and the cost shared between Sidbec and Norforge. The loading hoppers were enlarged at each mill feed point and consequently double handling has been virtually eliminated. Norforge delivers our grinding balls in these bins and collects the empties.

We have been operating in this manner for 6 months and are very pleased with the saving in labour, double handling and overhead crane usage.

Our overhead crane in grinding has a 40 ton and a 10 ton hook. The new ball delivery bins have a net capacity of 10 tons of balls, however, this is at present being limited to an all up weight, including container, of less than 10 tons. The reason for this is that if the container weighs more than 10 tons gross, the 40 ton hook must be used and its raise and lower speeds are too slow for this application. We are, consequently, upgrading our 10 ton capacity hook to approximately 12 tons to overcome this restriction.

Usage of balls to date has been approximately 3.4 kg/ton (7.5 lbs/ton) which is higher than originally estimated. Differences in performance of the Norforge, Burlington or Q.I.F. products have not become apparent yet as we have only achieved steady design production during the past few months.

The process water used is acidic with a PH of 5.5 and it is possible that the corrosion caused by this is contributing to our high grinding ball consumption. Consequently, during the past few months, we have been adding limestone to our process water in order to bring the PH to an alkaline 7.5 PH.

The effects of this have not as yet become noticeable.

F - Mill Bearings

The ball mill is supported on two hydrostatic trunnion bearings which consist of replaceable bronze liners in a cast Meehanite shell spherically seated in a cast Meehanite base and enclosed by fabricated steel covers. The necessary lift and thrust are supplied by a hydraulic unit located again under the mills, unfortunately in the area where the small pump boxes also overflow.

The trunnion bearing is provided with two lift points, each supplied with a minimum of 60 litres/min (16 USGPM) at 5600 KPA (390 PSI) from the hydraulic unit. The discharge end bearing is also provided with two thrust points having minimum requirements each of 19 litres/min (5 USGPM) at 2200 KPA (150 PSI). The combination fluid flowrate indicators cum switches are of the all mechanical type as supplied by Universal Filters, which are quite satisfactory for this application.

The minimum requirements mentioned here are also the trip points to stop the mill by clutch disengagement. As mentioned before, failure of the clutch to disengage will now result in motor stoppage and complete mill shutdown after 10 seconds.

The main purpose of this interlock is, of course, to save the bearings. Each mill has a total of 28 interlocks which will disengage the clutch when activated.

Of these 28 interlocks, 6 are flowmeters for lift or thrust, 8 are pressure switches and 14 are temperature switches.

During our start-up year of 1978, we have changed out 3 bearings, 1 feed and 1 discharge on mill "A" and the discharge end bearing on mill "C".

The exact dates and causes of failure have not been defined with certainty as the first failure only became apparent as a result of the discovery of bronze particules in the oil of mill "A" and subsequent efforts to trace an oil leak. In this case, the oil was in fact leaking through a crack in the bearing at the connecting point of the hydraulic lift.

It was deduced that this bearing failure had been caused by a malfunction of the solenoid valve when it had failed to open when called upon to do so, as a result of hydraulic pump stoppage. It is assumed that the damage to the bearing had occurred in the interval between the clutch failing to disengage and the arrival of the foreman on the scene who had the presence of mind to stop the motor.

A similar event had occurred on mill "C" and consequently its bearings were inspected, although no evidence of any problem was apparent on this mill. Slight damage to the discharge end bearing was discovered so it was changed rather than risking damaging the trunnion sometime in the future.

Following these bearing change outs, all solenoid valves were changed, a program of regular cleaning and lubrication of solenoids was commenced and most important, the interlock to stop the motor was fitted on all mills.

As a result of these modifications, no more problems have been experienced in this area.

G - Hydraulic Units

Six of these units are located on the ground floor, one under each mill. This allows for gravity return of the oil from the bearings to the tank.

The hydraulic unit consists mainly of a 3300 litres (900 US gal) tank with one high pressure Allweiler 380 litres/min (100 USGPM) pump to supply the lift and thrust for the bearings via a flow divider. A low pressure circulating De Laval IMO pump 510 litres/min (135 USGPM) is used to condition the return oil via strainers, 40 micron Parker Hannifin duplex filters and a heat exchanger. Magnetic centres have recently been installed in all these filters.

Our major problems in this area have been caused by oil contamination with iron ore concentrate resulting in the necessity to change 2 high pressure pumps, 2 circulating pumps and 1 flow divider during 1978. The sources of contamination are as follows: the feed end bearing is protected by a labyrinth seal which is located adjacent to the feed chute. Feed chute blockages during start-up were very common which resulted in changing out all feed chutes to larger diameter throat units. However, these blockages resulted in concentrate spilling over the feed end bearing cover which is protected by a labyrinth seal that is only partly effective. Probes to stop feed in the event of chute blockage have been installed in order to eliminate this spillage and the labyrinth seals modified by means of an extra cover. It was also suspected that there was insufficient bolting of the seal flange to the trunnion and the number of bolts on all these flanges was doubled.

At the discharge end, a considerable amount of steam flows over the labyrinth seal and gear cover, and deposits a film of ground ore on both. Discharge slurry temperature is approximately 65°C (150°F). Tests with a fine water spray to condense the steam resulted in detrimental dilution of the discharge slurry. A stack has been erected over mill "B" trommel screen and is quite effective. Upon perfection of this design to duct the maximum possible amount of steam away from the bearing and gear guard, they will be installed on all mills.

The other main source of contamination is caused by the location of the hydraulic unit. Frequent hosing of the floor in this area results in a thick film of slurry covering the tank, pumps, motors, filters, etc.

H - Feed Chutes

The original Nihard feed chutes at 250 mm (10 in) throat proved to be too small for the application and were all changed to a larger throat diameter of 430 mm (17 in). One chute has recently worn out and has been changed. A new chute is being fabricated in the shop which will have the facility of allowing its liner plates to be changed out rather than having to change out the complete chute.

The area of maximum wear is also the point where balls or slugs impact when the mill is being charged. We are working on a modification to introduce the balls at a more tangential angle in the chute to overcome this.

SUMMARY

Mill "D" has been taken as a test unit and its hydraulic unit is being raised off the floor by 2 metres (6 feet) which is the maximum possible height consistent with gravity return of oil to the tank. A cover will also be installed over the hydraulic unit.

At the same time, the unit will be modified by the installation of two higher capacity IMO high pressure pumps, one operating and one standby. A flow control valve, manually adjustable with bleed off to the tank, and a standby circulating pump will also be installed.

The purpose of this is to obtain the maximum possible availability from the hydraulic unit which we anticipate will determine the availability of the mill itself.

The electrical control system supplied with the mills is of the hardwire relay type.

As the rest of the plant is under the supervision of a GE logitrol programmable controller system, we anticipate incorporating all mill circuits within this system.

Apart from the advantages of eliminating all the relays, this programmable control system will give us the ability to troubleshoot the mill circuits via a monitor located beside the processing unit. This system also provides, among others, facilities for overriding of any contact, and for programming timers or counters into any circuit via the monitor keyboard.

When steady design production of 6 million tons per year from our two pelletizing machines is achieved, the one limiting factor to higher tonnages will be the throughput of ground ore.

Consequently, our maintenance resources at Sidbec-Normines will be directed in this area during this year in order to increase and maintain high availability of our ball mills.

Discussion of Paper by J. Wallace

P. Capper: What is your present availability in the ball mills ?

Reply: We have six mills available for one machine. If one mill becomes unavailable we start up another. All of the mills discharges are interconnected to a common header. Our availability has been about 80% but we intend to bring it up to 95%.

K. Konigsmann: Have you had any blocks on your pinion bearings ?

Reply: The helical configuration developes an axial thrust. The discharge end bearing has two thrust points to offset the thrust developed by the pinion. We have burned out one clutch but it was because of an operator problem.

P. Capper: You have two loop belts, one on concentrates and the other on pellets. How are they making out as to maintenance and operation ?

Reply: We have one 500 tph Stevensen-Adams loop belt feeding the up-grading plant. The other is a 7200 tph belt used in the ship loading. Both of them have operated very satisfactorily. The ship loading one has had a piece of metal break one of the idlers. Other than that we have not had any problems.

L. Dietrich: Have you considered any other reagent in the ball mill that will reduce the corrosion, if the high steel consumption is caused by corrosion ?

Reply: The only alternative considered so far is to bring the pH up by adding dolomite, which we are now doing. We do not know if it is successful.

K. Koningsmann: What is the volume of steel balls in the mill ?

Reply: The 400 tons of balls takes up 35% of the volume.

Paper No. 8

pH Control of
Mattagami Lake Mines and Orchan Mines Effluents

by

K. Stowe
Assistant Mill Superintendent

and

G. Veillette
Mill Superintendent

Mattagami Lake Mines Limited

Presented at Canadian Mineral Processors
January 23-25, 1979

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TABLE 1

CHIEF MINERAL CONSTITUENTS

MINERAL	MATTAGAMI	ORCHAN
PYRITE	25	30
PYRRHOTITE	10	20
SPHALERITE	15	15
CHALCOPYRITE	2	3
TOTAL SULPHIDES	52	68

MAGNETITE	13
VARIOUS SILICATES AND CARBONATES	35

pH Control of Mattagami Lake Mines and Orchan Mines Effluents

1. Introduction

Mattagami Lake Mines and Orchan Mines operate zinc-copper mines adjacent to one another near Matagami, Quebec, situated approximately 300 kilometres north-east of Rouyn-Noranda. (Note: Mattagami Lake and Orchan are in the process of being absorbed by Noranda Mines and will become Noranda Mines-Matagami Division). Both concentrators began operations in late 1963.

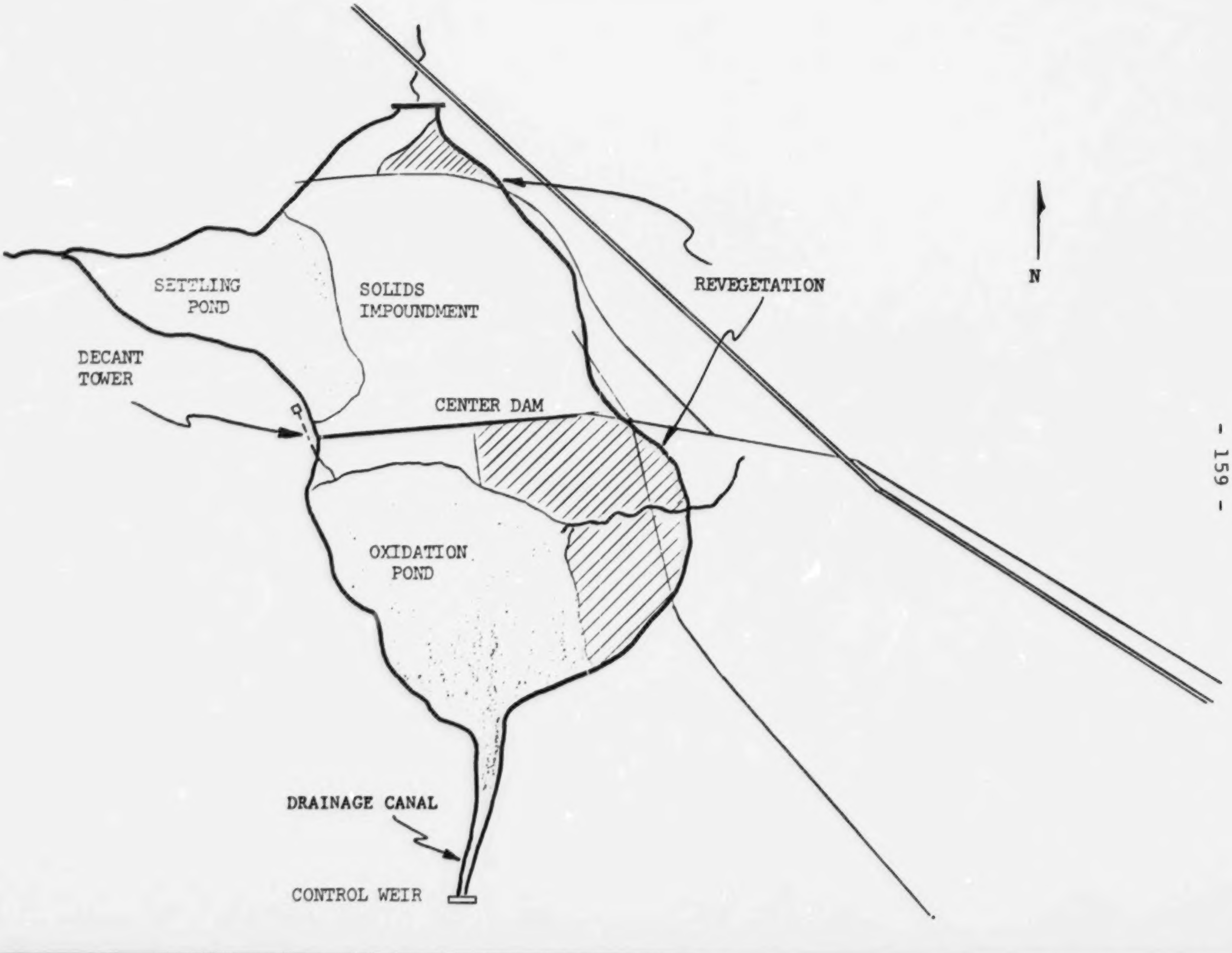
The Mattagami Lake concentrator is capable of treating up to 3,500 metric tons per day of massive sulphide ore with a mineral composition shown in TABLE 1. The Orchan concentrator consists of two separate circuits - the Orchan circuit with a capacity of 1,050 metric tons per day and a custom circuit with a capacity of 825 metric tons per day. Throughout its operation, Orchan has custom milled ore from several smaller deposits in the Matagami district.

2. Tailings Disposal

Mattagami Lake and Orchan pump classified tailings (at both mines backfill is recovered for filling underground) separately

FIGURE 1

WATSON TAILINGS POND


N
REVEGETATION
SETTLING
POND
SOLIDS
IMPOUNDMENT
DECANT
TOWER
CENTER DAM
OXIDATION
POND
DRAINAGE CANAL
CONTROL WEIR

to a common disposal area some 5 kiolmetres to the north-west. The disposal site was originally a shallow lake, with an average depth of 1 metre, flooding approximately 200 hectares and discharging in the Bell River. In order to prevent contamination of this river, from which the town of Matagami draws its water supply, a drainage canal was dug to divert the tailings effluent into the nearby Allard River (FIGURE 1).

With the space available for tailings disposal becoming restricted in 1970, and with the need to meet future government regulations and effluent quality standards, a long range program for upgrading the tailings disposal area was embarked upon. This program consisted of:-

1) construction of dams to isolate the tailings area from the surrounding countryside while providing sufficient volume for tailings disposal for the life of the mines. The bulk of the dam construction was to be done by spigotting

2) creation of a solids containment area and effluent oxidation pond by construction of a dam across the centre of the tailings area. This would hopefully minimize contact between acid effluents and the tailings thus reducing the dissolution of metals

3) revegetation of exposed tailings accessible to equipment

4) monitoring of the effluent and receiving streams on a regular basis.

Environmental costs were to be borne by the two mines in proportion to the quantities of tailings deposited.

3. Dam Construction

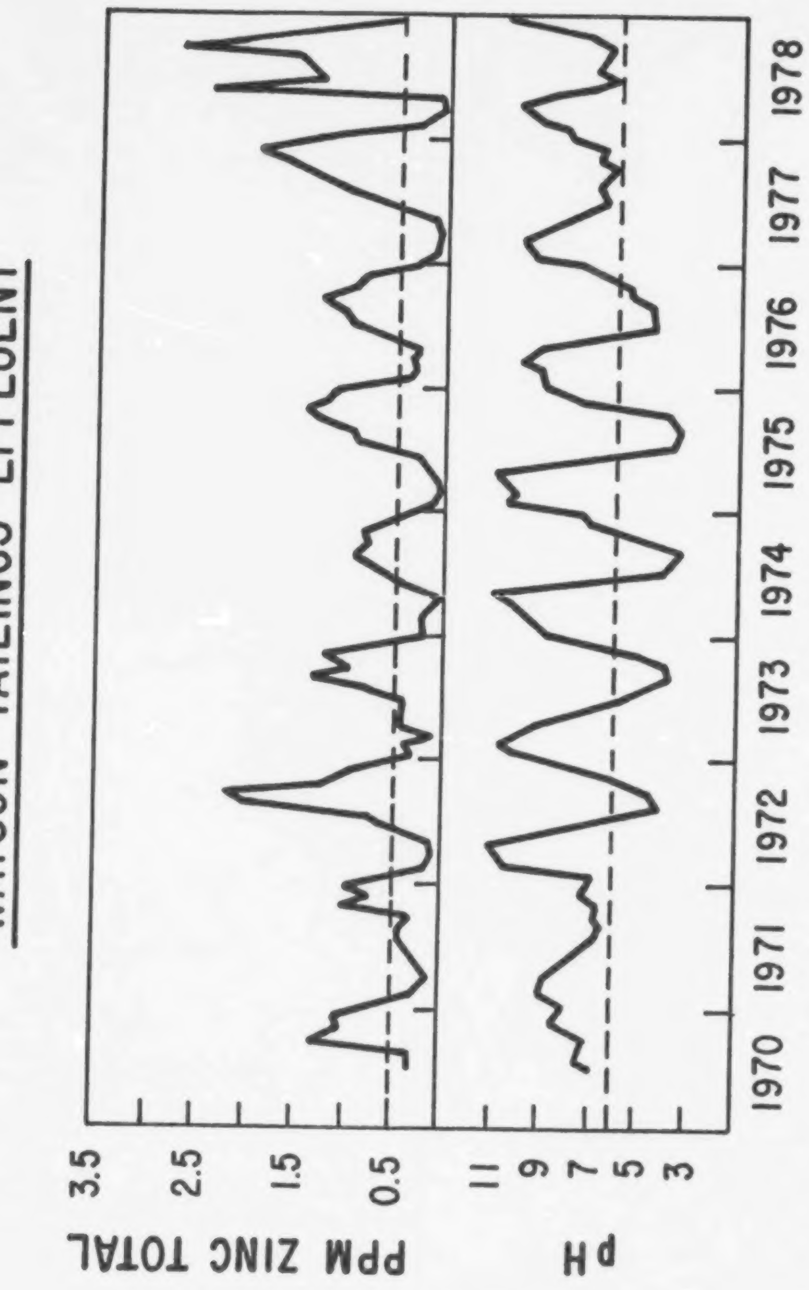
In 1970, a control weir was built across the drainage canal to allow for the adjustment of the pond retention time. This also provided a means of determining the quantity of effluent being discharged.

In 1972, the tailings dump points were moved to the northern end of the pond so that construction of dams enclosing the tailings area could begin. Two years later construction of the dam dividing the pond into two sections was started. During 1976, the solids impoundment area and effluent oxidation pond became a reality. 1977 saw the completion of a decant tower to transfer clear overflow from one section of the pond to the other.

In the spring of 1977, a diversion ditch was excavated and a clay dam erected to the northwest to prevent drainage into the pond from that area. In the future, drainage into the pond from the southeast may be similarly diverted.

FIGURE 2

WATSON TAILINGS EFFLUENT



4. Pond Acidification

Upon completion of the control weir, the quality and quantity of effluent began to be monitored on a regular basis - weekly during the summer, bi-weekly during the winter. The receiving stream, the Allard River, also began to be analyzed for dissolved metals both upstream and downstream of the effluent discharge.

In the summer of 1972, the tailings effluent acidified rapidly after break-up, the pH of the entire pond eventually falling below 4. It remained thus until the onset of winter. An identical pattern was experienced in subsequent years (FIGURE 2). Co-incident with the drop in pH, the levels of dissolved metals, most notably zinc, iron and manganese, increased markedly.

The biological decompostion of thiosalts during the summer was identified to be the source of the acid generation. However, the question still remained why acidification had not occurred in 1970 and 1971. Two major changes were made during 1972 which were thought to provide possible explanations:-

1) sulphur dioxide had begun to be used in the Mattagami Lake concentrator to improve copper concentrate grade

2) moving of the tailings dump points to the northern section of

IMPORTANT THIOSALT REACTIONS

FORMATION

$$FeS_2 + 2\,OH^- \longrightarrow FeS + S^{--} + H_2O + \frac{1}{2}O_2$$

$$Fe_7S_8 + 2\,OH^- \longrightarrow 7\,FeS + S^{--} + H_2O + \frac{1}{2}O_2$$

$$2\,S^{--} + 2\,O_2 + H_2O \longrightarrow S_2O_3^{--}\text{(THIOSULPHATE)} + 2\,OH^-$$

(SIMILAR REACTIONS PRODUCE POLYTHIONATES S_nO_6, where $n = 3,4,5,6$)

$$2\,S_2O_3^{--} + 3\,SO_2 \longrightarrow 2\,S_3O_6^{--}\text{ (TRITHIONATE) } + S$$

GENERATION OF ACID

$$S_2O_3^{--} + 2\,O_2 + H_2O \longrightarrow 2\,H^+ + 2\,SO_4^{--}$$

$$S_3O_6^{--} + 2\,O_2 + 2\,H_2O \longrightarrow 4\,H^+ + 3\,SO_4^{--}$$

NEUTRALIZATION WITH LIME

$$S_2O_3^{--} + \frac{3}{2}O_2 + 2\,CaO \longrightarrow 2\,CaSO_4$$

TABLE 2

the pond to commence dam building had greatly increased the effluent retention time, thus allowing for more complete oxidation.

5. Thiosalt Generation

Processing of sulphide ores by conventional methods results in the generation of thiosalts - thiosulphate, $S_2O_3^=$, and polythionates, $S_nO_6^=$ where n = 3,4,5,6. During grinding, aeration and flotation, a small amount of sulphur from the sulphide minerals dissolves in the process water (TABLE 2). The dissolved sulphur then oxidizes to form thiosulphate and the various polythionates. Polythionates, trithionate $S_3O_6^=$ in particular, can also be produced by the reduction of thiosulphate by sulphur dioxide.

Thiosalts that are generated during processing can then oxidize in the tailings pond to produce acid. Although the rate of oxidation is influenced by several factors, the most critical in this case is temperature due to the presence of bacteria in the tailings pond which act as a catalyst for the oxidation reaction. At colder water temperatures,bacterial activity is greatly diminished. In the winter months the thiosalt oxidation rate becomes almost negligible and insufficient acid is generated to cause a drop in the pond pH. During the summer, on the other

TABLE 3

CONCENTRATOR THIOSALT PRODUCTION

GRINDING CIRCUIT

ROD MILL	15 %
No. 1 & 2 BALL MILLS	15
No. 1,2 & 3 AERATORS	15
	45

COPPER FLOTATION

ROUGHING, SCAVENGING	15 %
CLEANING (SO_2 ADDITION*)	10
	25

ZINC FLOTATION

ROUGHING, SCAVENGING	25 % **
CLEANING	5
	30

** SULPHUR DIOXIDE CONSUMPTION 0.3 lb/ton*

***THOUGHT TO BE DUE PRIMARILY TO OXIDATION OF SULPHUR DIOXIDE TO THIOSALTS DOWNSTREAM FROM POINT OF ADDITION.*

hand, virtually all of the thiosalts oxidize before leaving the pond thereby producing an acid effluent. It is estimated that a pond retention time of two weeks is required for complete oxidation of the thiosalts during the summer.

An estimate sampling program of the Mattagami Lake concentrator was carried out during 1976 to determine the main sources of thiosalts and in particular the contribution due to the use of sulphur dioxide. Sulphur dioxide is added during copper cleaner flotation (the pH is lowered to 4.5) to aid depression of sphalerite and pyrite and give a significant increase in copper concentrate grade.

The results of the sampling program are given in TABLE 3. Grinding and aeration were found to be major sources of thiosalts. An estimated fourty percent of the thiosalt generation was found to be attributable to the use of sulphur dioxide. Furthermore, much of the sulphur dioxide was found to oxidize to thiosalts during zinc rougher flotation. This would be reasonable since thiosalt formation is favoured under alkaline conditions and not the acidic conditions which prevail during copper cleaning.

Even without the use of sulphur dioxide, Mattagami Lake's tailings would still contain a considerable amount of thiosalts - indeed, more than enough to cause an acidic effluent given adequate retention time for thiosalt oxidation. Monitoring of

TABLE 4

THIOSALT* CONTENT OF MILL TAILINGS

YEAR	MATTAGAMI LAKE		ORCHAN	
	No. OF SAMPLES	AVERAGE PPM $S_2O_3^{--}$	No. OF SAMPLES	AVERAGE PPM $S_2O_3^{--}$
1976	16	143	6	153
1977	20	131	16	132
1978	15	128	13	122
1976-1978	51	134	35	132

* *ANALYZED AS THIOSULPHATE*, $S_2O_3^{--}$

Orchan's (which does not use sulphur dioxide) tailings since 1976 has shown it to contain levels of thiosalts almost identical to Mattagami Lake's (TABLE 4). The onset of tailings pond acidification in 1972 would therefore seem to be due to the change in retention time rather than the use of sulphur dioxide.

6. Effluent Neutralization

An experimental program was initiated in 1976 to determine if acidity produced by thiosalt oxidation during the summer could be neutralized by the addition of hydrated lime to the tailings pump box at the Mattagami Lake concentrator. A large savings in capital cost, labour, maintenance and supervision could be realized if lime addition at the concentrator was found to be effective due to the following reasons:-

1) the control weir and decant, the best two locations for lime addition, are virtually inaccessible during the summer months, except on foot.

2) large quantities of lime - up to 150 tons per month - would have to be transported from the unloading area at the concentrator to the tailings area.

3) unloading, storage and feeding facilities would be required at the tailings area

4) power would have to be supplied

5) supervision of the site would be limited due to the distance from the minesite.

In addition, a possible improvement in effluent quality appeared likely. Since the tailings water would never be allowed to acidify the dissolution of metals was expected to decrease.

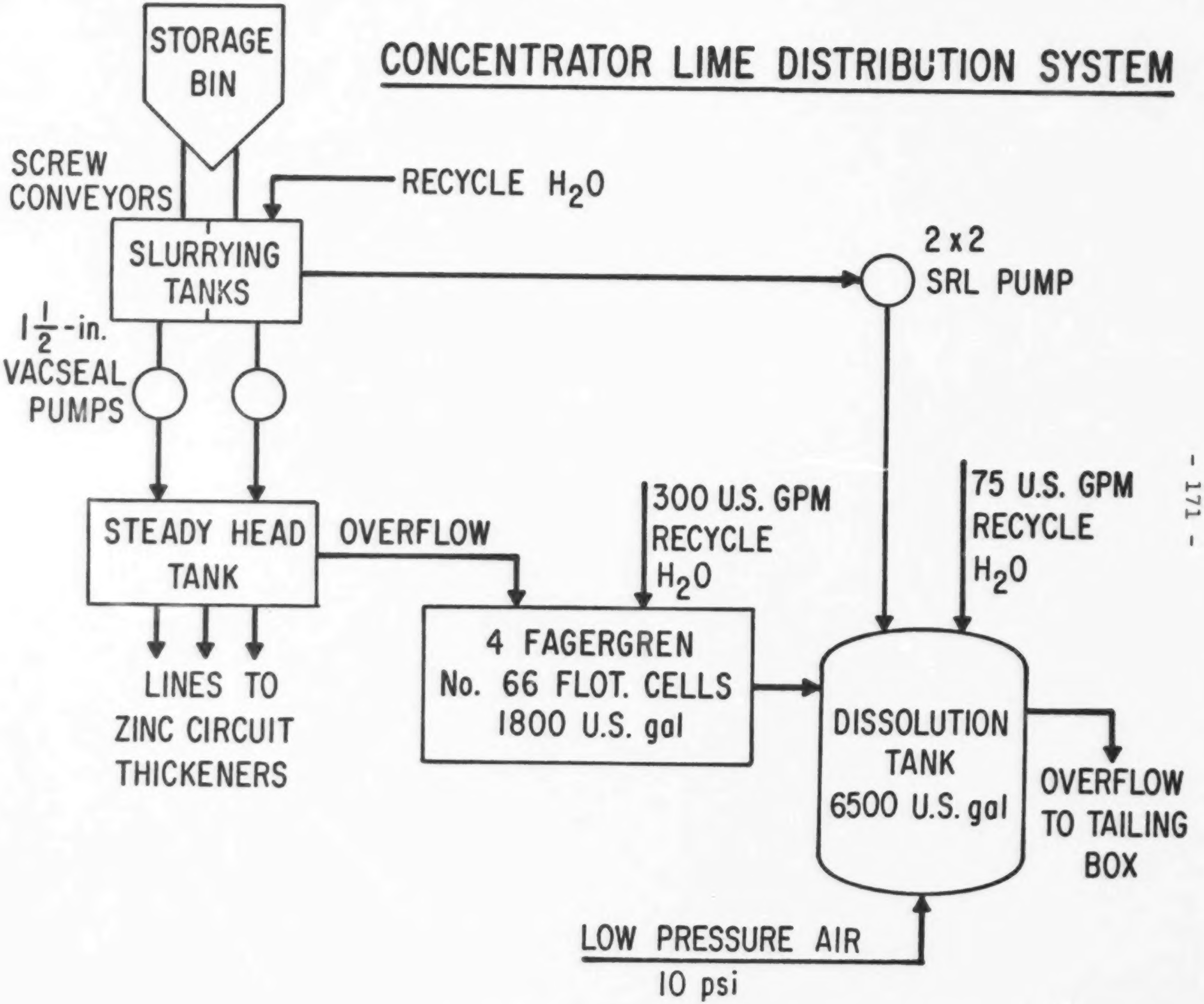
Initially, lime was pumped directly to the tailing box from the existing lime slurry tanks. This led to an unexpected problem as hydrated lime added in solid form to the tailings did not subsequently dissolve. Instead, it somehow became entrained in the tailings and the bulk of the lime (more than seventy percent), being unavailable for neutralization, was wasted.

To correct this situation, it was decided to dissolve as much of the lime as possible prior to its addition to the tailings. An unused reagent tank, capable of providing several minutes retention time, was conveniently located beside the tailings box. Several hundred gallons per minute of water pumped from underground, which had previously been discharged unused, was also readily available.

When this system became operational near the end of the summer of 1976, an eighty percent dissolution rate of lime was achieved. Sufficient dissolved lime was available to completely neutralize the thiosalts present in the tailings, as samples of

FIGURE 3

CONCENTRATOR LIME DISTRIBUTION SYSTEM


STORAGE BIN
SCREW CONVEYORS
RECYCLE H_2O
SLURRYING TANKS
2 x 2 SRL PUMP
$1\frac{1}{2}$-in. VACSEAL PUMPS
STEADY HEAD TANK
OVERFLOW
300 U.S. GPM RECYCLE H_2O
75 U.S. GPM RECYCLE H_2O
LINES TO ZINC CIRCUIT THICKENERS
4 FAGERGREN No. 66 FLOT. CELLS 1800 U.S. gal
DISSOLUTION TANK 6500 U.S. gal
OVERFLOW TO TAILING BOX
LOW PRESSURE AIR
10 psi

tailings water collected at this time and allowed to stand, stabilized at pH 7 when acidification was complete.

The lime dissolution system was modified during 1977 and 1978 to improve its operation and provide further dissolution capacity. The present system can be seen in FIGURE 3. It has proven to be extremely reliable with little or no supervision required.

Approximately two-thirds of the total lime used is pumped from the slurrying tanks to a steady head tank, from which lime is distributed to the various other points of use in the mill. The remaining one-third of the lime is pumped directly to the dissolution tank. The overflow from the steady head tank goes to a bank of four #66 Fagergren flotation cells where 300 gallons (U.S.) per minute of underground water is added. The discharge of the cells flows to the dissolution tank where additional water is added. The tank is agitated by low pressure air and must be drained periodically of accumulated insoluble material.

7. Effluent Quality

The effluent was maintained above the target level of pH 6 throughout 1977 and 1978. The overflow from the solids impoundment area to the oxidation pond was extremely low in

TABLE 5

AVERAGE SETTLING POND AND FINAL EFFLUENT ANALYSES

(June-Sept. 1978)

	pH		PARTS PER MILLION							
			Zn	Cu	Fe	Mn	CaO	SO_4^{--}	$S_2O_3^{--}$	SUSP SOLIDS
SETTLING POND EFFLUENT	9.1	TOTAL	0.13	0.06	0.94	0.10	145	266	46	10
		DISSOLVED	0.06	0.03	0.26	0.06	133			
FINAL EFFLUENT	6.5	TOTAL	1.94	0.10	1.43	0.87	123	278	2	12
		DISSOLVED	1.73	0.06	0.60	0.83	120			

suspended solids and dissolved metals (TABLE 5). However, the dissolved metal content of the final pond effluent was unexpectedly high at times.

It was noticed that the level of dissolved metals increased sharply after periods of heavy rainfall. Run-off from older, exposed tailings was found to be the source of dissolved metals. Grab samples of small flows of surface run-off have contained up to 50 ppm of dissolved metal. Methods of preventing or reducing this leaching action are currently being investigated.

Discussion of Paper by G. Veillette and K. Stowe

F. Makarinsky:	Have you found any build-up of scale on your overflow pipe from the tailings pond ?
Reply:	No, not at all. It has been running three summers now. The overflow pipe is about 8 or 10 inches in diameter.
E. Joe:	How long do your electrodes last in the pH control unit ?
Reply:	We do not use tne unit on a continuous control basis but take regular grab samples.
Question:	Are you concerned with retention time in the treatment tank ?
Reply:	Yes. The more retention time the better the dissolution. Our time is about 10 to 15 minutes.

G. Johnston: Regarding weir control, do you find the dissolved mineral content highest when the flow rate is highest ?

Reply: Yes. When the weir is highest the minerals dissolve well. This is usually when it is raining and the flowrate is high. In spring it is 10 times normal. There is an 8 or 10 inch board to put in or take out for easy control.

W. Meekel: Do you have any plans for correcting the problem of high lime ?

Reply: Yes, but they are only ideas yet. This is the area that we will be concentrating on. We may have to put our pH control right at the lake to precipitate the dissolved metals.

W. Petruk: Can you give us some idea of the amount in the river ?

Reply: The upstream and downstream values are high compared to the effluent. There is only a slight change caused by the effluent. i.e. 0.01 to 0.012 ppm zinc.

W . Meekel: Your tailings now appear to be quite acid probably from the oxidation.

Reply: The tailings to start are not acid. In some places now where the grass has been growing some 4 or 5 years, it is starting to die because it is turning acid. The best growth areas are not yet acid. There might be some segregation of particle size or mineral.

E . Rolia: Do you find a change in pH in the discharge from your oxidation lake ?

Reply: No. The flow is small compared to the river flow.

IMPORTANT CONSIDERATIONS
IN
CYANIDATION OF GOLD ORES

by

E.W. Johnson

The current increase in gold price may presage a resurgence of gold milling plants in Canada similar to that of the thirties. Because the industry has been at a low ebb for almost two decades, future operating responsibilities will likely devolve upon personnel with limited cyanidation experience. My comments, which may seem elementary to experienced operators, are directed towards these industry newcomers to highlight operational features in cyanidation which forty years' experience has shown me to be of importance. A brief review of what actually takes place in a cyanide circuit may help give a better understanding of the topics to be discussed. These will include: (1) expose the gold
(2) dissolve the gold
(3) make solid/liquid separation
(4) recover the gold from solution.

1. Expose the Gold

The chemical key to gold recovery by cyanidation lies in the ready solubility of the metal in a relatively weak solution (0.04%-0.075% NaCN equivalent by weight) of calcium or sodium cyanide. Obviously, dissolution cannot start unless the cyanide is in contact with the gold. As an analogy, the yolk of a hard-boiled egg can be taken to represent the gold particle and the egg white and shell, the surrounding gangue or sulphide host. As the shell and white must be fractured to expose the yolk, similarly the surrounding gangue or sulphide host must be broken or fractured to enable the cyanide solution to contact the gold.

The major dimension of the gold will range in very rare instances from 15 - 20 plus centimeters to the more common 200 - 300 microns and down to fractions of a micron. The coarse gold is readily unlocked in crushing and primary grinding and recovered in jigs. The ultra-fine gold is not always amenable to release by fine grinding in which case roasting or smelting must be evaluated.

In cyanidation the grinding requirement is to unlock coarse gold particles (+100 mesh) to free them for recovery by jigs or traps and to unlock (but not necessarily free) finer particles to expose them to contact with cyanide solution.

Usual plant grinding practice includes one or more of the following:

(i) jigging to concentrate and recover mid-to-coarse gold. This gold is removed from the circuit and requires only refining prior to marketing. Recovery range 5 - 75%.

(ii) grinding in cyanide solution - if the ore is "clean" quartz-pyrite and free of cyanicides - recovery up to 80% in grinding circuit.

(iii) grinding in water - if ore contains soluble copper or unstable minerals such as pyrrhotite.

(iv) grinding in the presence of lime - if ore contains pyrrhotite.

2. Dissolve the Gold

The gold is dissolved by aerated cyanide solution in agitation tanks. These agitators are designed to keep the ore particles in suspension while allowing maximum contact between the gold particles and oxygenated solution. There are three common types:

(i) Dorr - rotating central air lift column - to which are attached two rakes at the bottom and two pulp distributing launders at the top.

(ii) Denver-Devereux - propeller below a draught tube imparts a rolling turbulence to pulp. Side air lifts are sometimes added to increase aeration.

(iii) Pachuca - conical bottom tank, ratio height to diameter about 6 to 1; central and side air lifts to aerate pulp and keep solids in suspension.

The number of agitators in a circuit will be governed by the required contact time for optimum gold extraction and the density (unit volume) of the feed pulp.

Mention should be made of three uncommon phenomena that can occur at this stage of the process:

(i) adsorption of gold by "active" graphite in the ore - can sometimes be inhibited by the addition of kerosene to the pulp, prior to the addition of cyanide.

(ii) re-precipitation of gold by the inter-action of highly alkaline, low strength cyanide on fine sulphide ore.

(iii) regeneration of cyanide in the agitation circuit, attributed by Hamilton to inter-action of soluble sulphides, double salts and free alkali. During such occurrences cyanide strength as indicated by titrations will rise to two or three times normal levels. Cyanide additions to the circuit are maintained at about one quarter normal until conditions return to normal.

During agitation, gold is dissolved and converted from solid to solution form.

3. Make Solid/Liquid Separation

The next process stage is to separate the gold bearing solution from the residual solids. In the past this has been done by counter current decantation in a series of thickeners or alternatively by single or multi-stage filtration or a combination of the two. The choice is a matter of economics. In cyanidation practice, drum type filters having the better washing efficiency, have been in more common use than have disc filters. Recent technology points towards belt type filters for increased capacity and better washing. Activated carbon technology is receiving increased attention as a method of recovering gold from solution and eliminating the need for thickeners, filters and a Merrill-Crowe precipitation plant.

4. Recover the Gold from Solution

The final recovery step is reversal of the dissolution operation in that the soluble gold compound is rendered solid, by being drawn out of solution by the addition of finely powdered metallic zinc. The process is the familiar Merrill-Crowe "instantaneous" clarification, de-aeration and precipitation process. In this system, operating under vacuum, the pregnant (gold bearing) solution is drawn through cotton-covered (10 oz. duck) pipe-frame filters to remove fine suspended solids. Good clarification is essential for complete precipitation (low grade barren assays) and low press pressure. The clarified solution is piped to a de-aerating tower for the removal of dissolved oxygen. The elimination of oxygen is critical in precipitation operation to prevent loss of zinc metal and re-dissolution of gold in the filter press. Zinc dust is fed as an emulsion to the suction side of the precipitation pump which draws de-aerated solution from the tower. The addition of a lead salt (lead nitrate) to the pregnant solution prior to precipitation enhances the efficiencies of the zinc precipitant which acts almost instantaneously to remove precious metals and copper and nickel from solution.

The precipitated mixture of gold, zinc, lead, silica and other base metals such as copper, iron and nickel is recovered in a filter press, usually "sock" or bag type for low solution tonnages or the plate and frame, for larger tonnages. The barren press effluent is re-cycled to plant use and/or discard. The precipitate is removed from the press on a weekly to monthly schedule and delivered to the refinery for smelting and marketing.

The above review of cyanidation operations has outlined the main process steps. There are considerations of ultimate economic interest, related to these steps, which are sometimes overlooked. They are therefore deemed important enough to note, particularly for the new operator. The discussion topics include:

(1) soluble gold losses in cyanidation
(2) completing the metallurgical balance
(3) mineral solubility and solution quality.

1. Soluble Gold Losses in Solution

In cyanide gold milling it is important to remember that moisture in filter cake residue or tailing thickener underflow contains dissolved gold which must be accounted for along with that in the solids, in metallurgical calculations. This is done by firstly determining the solution/solids ratio on the tailings sample, filtering the pulp, then washing and drying the filter cake and finally, assaying the solution and solids.

The assay values are combined in the pre-determined ratio to obtain the total tailings loss. If an ore is concentrated prior to cyanidation, the losses will be factored by the ratio of the concentration, to determine the loss per ton of ore. The variance in loss between solid tails and total tails points up the difference between gold extraction and recovery in cyanidation.

Loss of soluble gold can be minimized in three ways:

(i) In gold milling, "recover the minerals as soon as they are free", - a principle widely publicized by the late A.C. Daman of Denver Equipment, is a good starting point. It is most often done by including jigs in the primary and/or secondary grinding circuits to recover plus 100 mesh metallics. Early recovery of gold lessens risk of later loss and as well, reduces solution grades in the cyanide circuit.

(ii) As a general rule direct high grade solutions to the head end of the leaching circuit and low grade solutions such as barren solutions, secondary filtrate and wash water to the final stages. Examples of this practice can be seen in almost all cyanide flow sheets.

(iii) Optimize filter operation so as to:

a) prevent filter feed pulp from overflowing the boot, into the tailings re-pulper,

b) maintain filter feed density at required level,

c) maintain optimum wash on filter cake. The wash can be either barren solution or water on the final stage depending upon the bleed requirements, solution quality or circuit solution balance,

d) provide adequate vacuum on filters - 22 - 25 inches mercury, or better - and maintain all valves, piping, filter cloths in a good state of repair.

2. Completing the Metallurgical Balance

This important consideration particularly emphasizes the need to monitor all sundry discharges from the mill. This is usually done by estimating flows and sampling for assay or alternatively by adding these discharges to the mill tailings prior to (and therefore to be included in) the final tailings sample. Examples of these sundry discharges include dewatering or desliming thickener overflows and/or barren solution or wash water, discarded to balance solution levels in the circuit.

All too often unmonitored flows containing "unconsidered trifles" make for embarrassing situations and leave metallurgical balance calculations less than complete.

3. Ore Solubility and Solution Quality

Gold cyanidation metallurgy embraces two important concepts. First, the solubility rates of the contained minerals and not the total metals content determine whether the ore is amenable to cyanidation.

Second, the response of an ore to cyanide treatment can ultimately be determined only by analysis of the mill solutions. The significance of these statements can be more fully realized by reading American Cyanamid Mineral Dressing Notes, No. 23, entitled "Chemistry of Cyanidation", published in December, 1968.

Operational control requires solution analyses for metal ions such as copper, nickel, zinc, arsenic and antimony, for guidance in adjusting barren bleed rates to maintain the concentration of critical elements within acceptable limits. These limits vary with different ores and seem to depend on interactions of constituents. The operating objective then becomes a matter of relating analytical data to plant performance and determining optimum conditions.

In many mills "reducing power" is important in that it reflects gold-dissolving efficiency of the cyanide solution. It is actually a measure of oxygen-consuming constituents usually comprised of iron and sulphur compounds derived from the dissociation of the iron sulphides:- pyrite, marcasite and pyrrhotite.

Reducing power is reported as volume (in millilitres) of one-tenth normal potassium permanganate solution (N/10 $KMnO_4$ = 3.16 grams $KMnO_4$ per litre solution), per litre of mill solution. In some mills reducing power above 500 is critical whereas in others the limit can exceed 1,000 - 1,500. In general, the lower the reducing power, the better the cyanide operation. Usually high reducing power (1,500 - 2,000+) is associated with reduced gold extraction and recovery, and attendant increased reagent consumption. In precipitation, high reducing power is characterized by high press pressure and poor barren assays.

Measures which have proven effective in lowering reducing power and improving solution quality include:

(i) pre-aeration of ore or concentrate in a lime pulp prior to cyanidation,

(ii) increased dilution of pulps in cyanide agitation circuit,

(iii) judicious discard of barren solution to control reducing power.

Conclusion

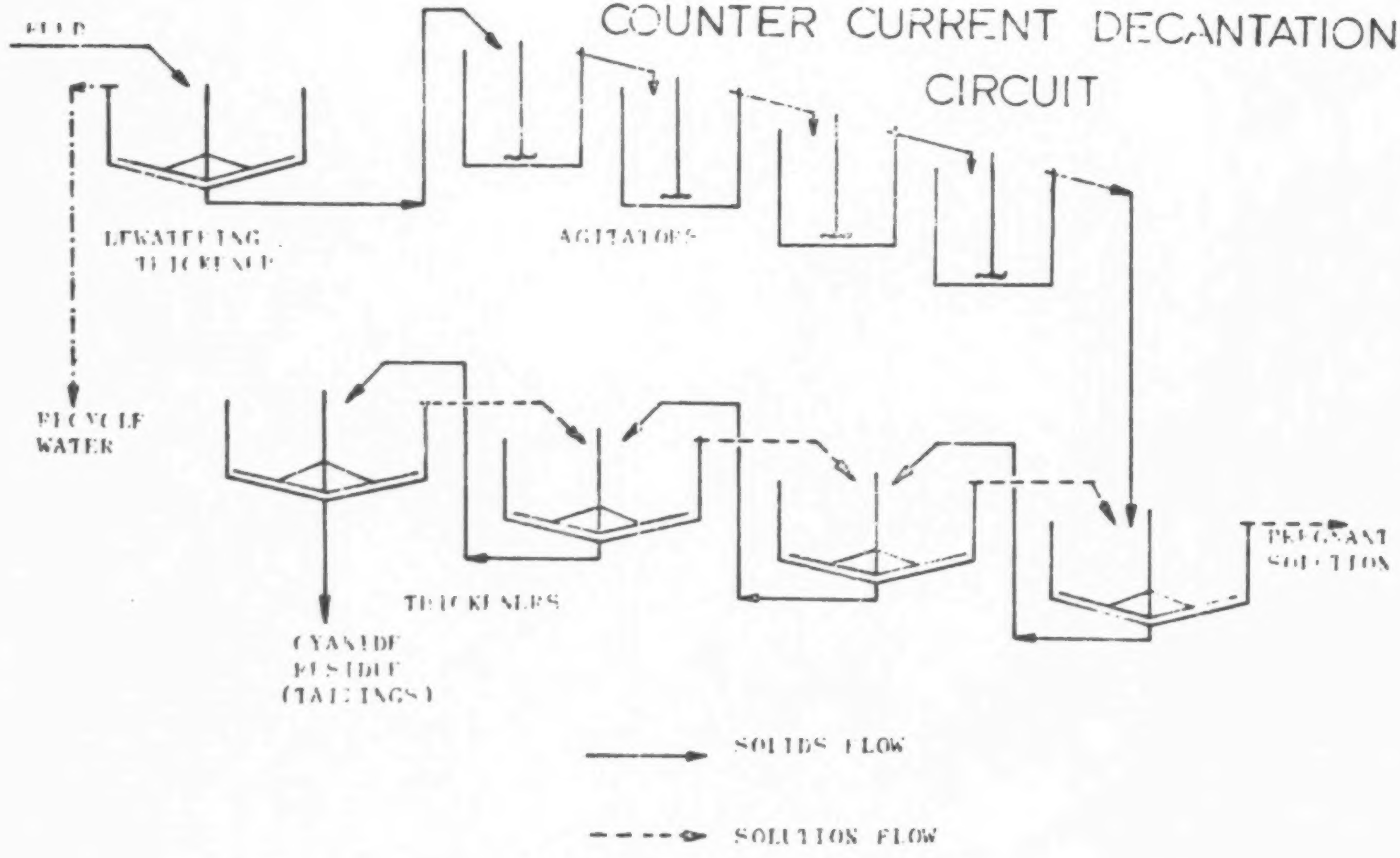
This paper will have served its purpose if it helps one cyanide operator become aware of the possible source of potential problems and motivates him to take corrective action. His operating objectives will be to minimize soluble gold losses; to make certain that an accounting is made of all plant losses; and to maintain solution working efficiency at best possible level.

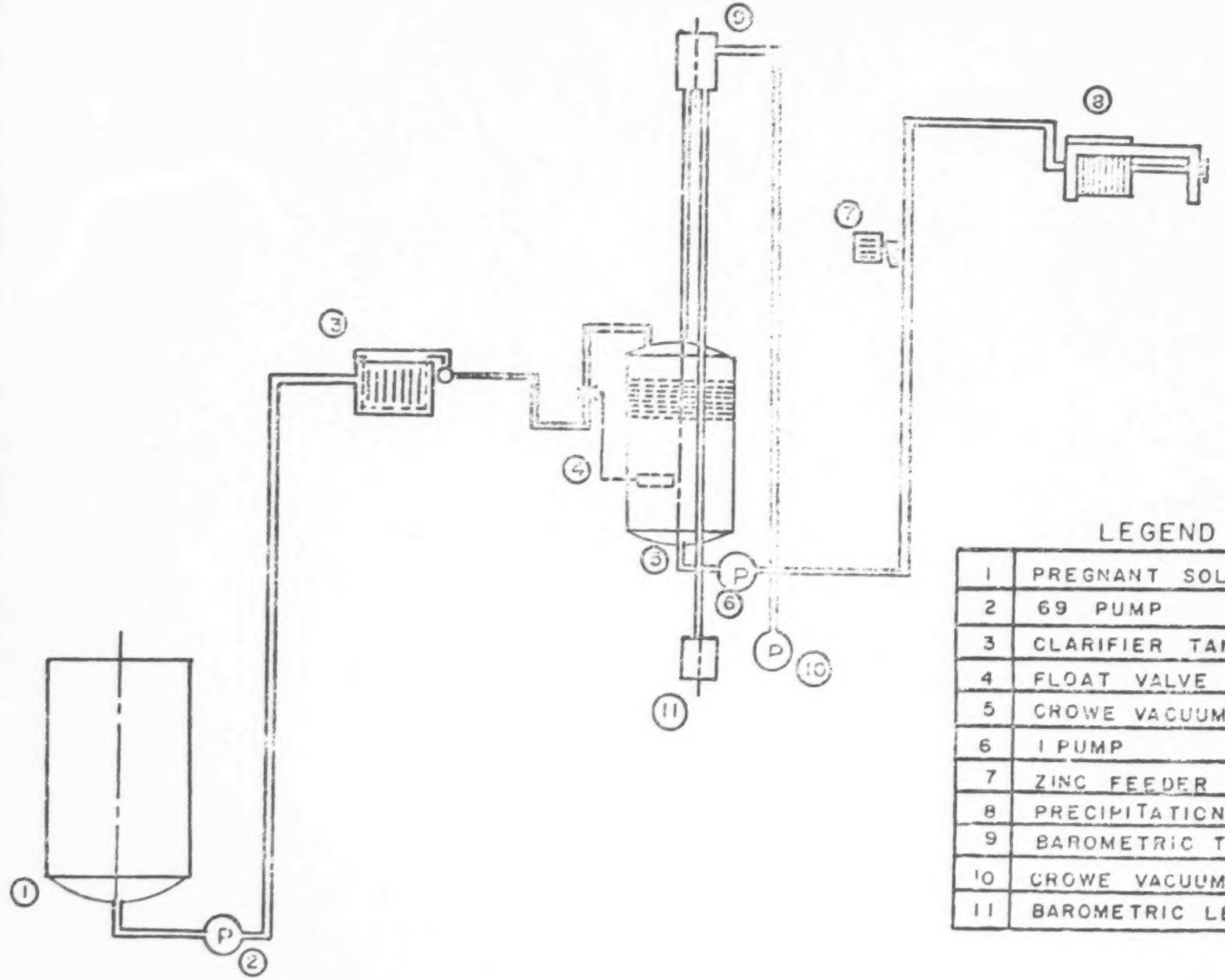
E. W. Johnson,
Supt. of Surface Crushing,
Milling and Laboratories.

EWJ:hm

15/01/79

COUNTER CURRENT DECANTATION CIRCUIT


FEED
DEWATERING THICKENER
AGITATORS
RECYCLE WATER
THICKENERS
CYANIDE RESIDUE (TAILINGS)
PREGNANT SOLUTION
SOLIDS FLOW
SOLUTION FLOW



LEGEND

1	PREGNANT SOLUTION TANK
2	69 PUMP
3	CLARIFIER TANKS
4	FLOAT VALVE
5	CROWE VACUUM TANK
6	1 PUMP
7	ZINC FEEDER
8	PRECIPITATION PRESSES
9	BAROMETRIC TANK
10	CROWE VACUUM PUMP
11	BAROMETRIC LEG TANK

PRECIPITATION CIRCUIT	
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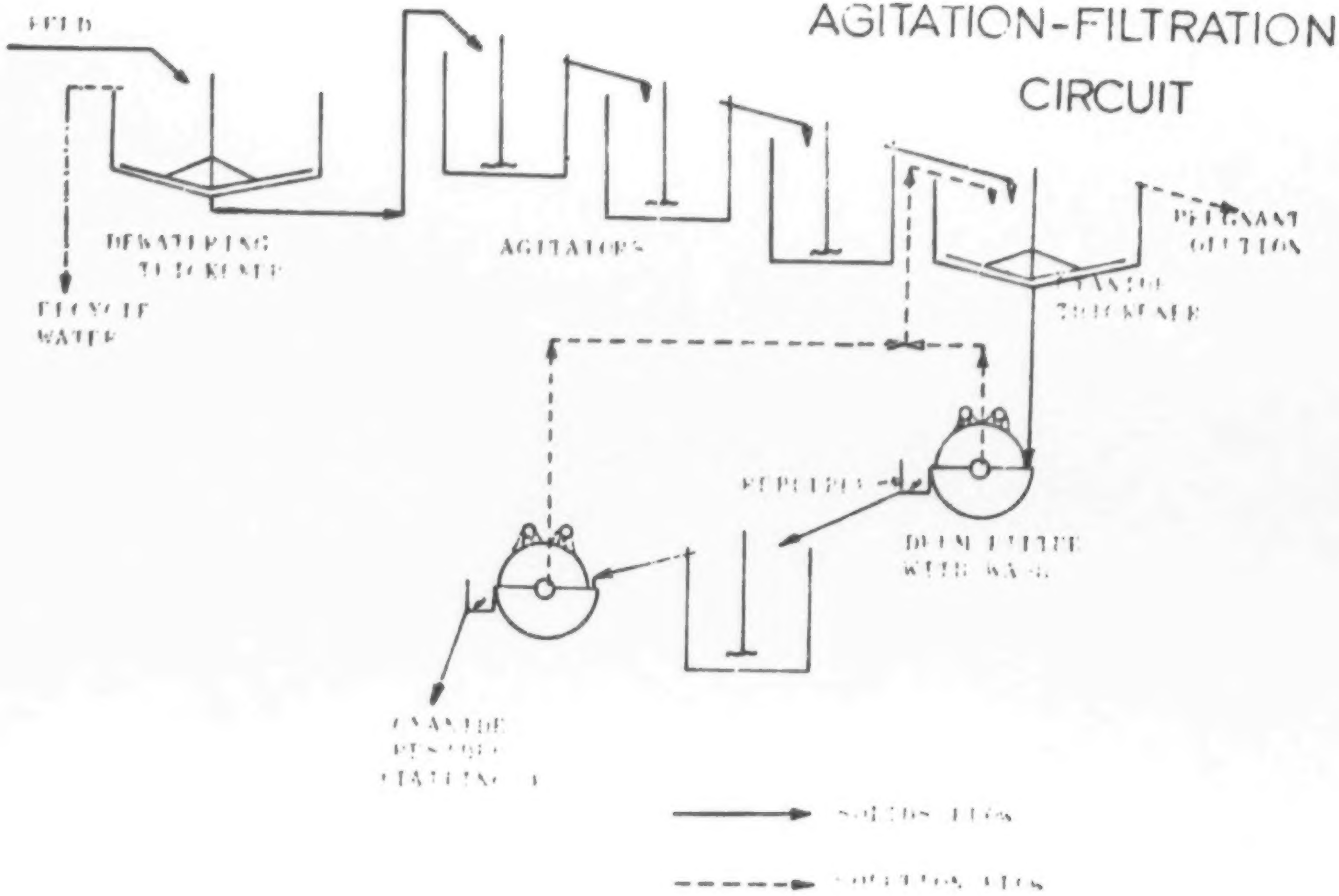
AGITATION-FILTRATION CIRCUIT


FEED
DEWATERING THICKENER
RECYCLE WATER
AGITATORS
PREGNANT SOLUTION
CYANIDE THICKENER
REPULPER
DRUM FILTER WITH WASH
CYANIDE RESIDUE STACKING
SOLIDS FLOW
SOLUTION FLOW

Discussion of Paper by E. Johnson

D. Livingstone: In bench testing of new ores are there any new techniques for anticipating the amount of barren that you would have to discard ?

Reply: You should analyze the solution for such things as copper and nickel. Keep in mind just how representative your sample is. Usually the only thing consistent is the inconsistency of the feed to the mill and the samples taken.

E. Marcotte: Have you determined if there would be any advantage of using carbon adsorption instead of the Merrill-Crowe precipitation method ?

Reply: No. There is probably no advantage in an old mill. If you were going to start a new mill it might be worth considering. We concentrate the sulphides and then regrind the concentrate before cyanidation. However, with the present high price of gold, we are considering cyaniding the flotation tailings. The carbon method of precipitation is being considered in order to avoid a major expansion of the present precipitation circuit.

Paper No. 10

Collectorless Flotation of Chalcopyrite and Sphalerite

by

R.H. Yoon* and A.I. Stemerowicz*

ABSTRACT

The flotation of chalcopyrite and sphalerite from copper and copper-zinc ores was achieved without using thiol-type collectors. The ore is first treated with sodium sulphide during grinding or the subsequent conditioning step, followed by flotation using frother only. For the copper ore the collectorless flotation results obtained were superior to those reported in a previous investigation using conventional flotation techniques. On the copper-zinc ore the best collectorless flotation results were obtained when the ground ore was aerated with sulphur dioxide prior to flotation. These results were comparable to those obtained in a comparison test using collectors.

It is believed that treatment with sulphur dioxide renders the mineral surfaces hydrophobic by removing surface oxidation products. The selective, collectorless flotation of chalcopyrite and sphalerite from other sulphides is explained by the differences in solubility products of the various sulphides.

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* Research Scientists, Mineral Sciences Laboratories, Canada, Centre for Mineral and Energy Technology, Department of Energy Mines and Resources , Ottawa, Canada.

INTRODUCTION

Sulphide minerals are readily amenable to flotation but not without proper surface treatment. Thiol-type collectors are most widely used to render selected constituents of a sulphide ore hydrophobic. However, at the Tsumeb concentrator in South Africa,galena and sphalerite are floated selectively from a chalcopyrite ore without using a collector in the first eight cells of each rougher bank.(1) Lepetic has also shown that chalcopyrite can be floated successfully by using a frother alone after dry, autogenous grinding.(2)

These recent reports on collectorless flotation have lead to a revival of the old controversy as to whether or not sulphide minerals are naturally hydrophobic. This controversy can be traced back to the days of oil and skin flotation, the principles of which are based on the fact that sulphide minerals are less wettable by water than siliceous gangue minerals. Since then, it was believed that many sulphide minerals were naturally hydrophobic, and Sulman and Gaudin noted that some sulphides such as galena and chalcopyrite were floatable without using a collector.(3) (4) Taggart and Wark opposed this view rather vigorously and contended that the apparent natural floatability was due to either contamination or by the use of frothers which had collecting properties.(5) (6) They also believed that clean specimens became contaminated and gave finite contact angles if they were permitted to dry in air. Plaksin offered a different explanation for the apparent hydrophobicity stating that it was due to the adsorption of molecular oxygen, which helps in promoting dehydration of sulphide minerals.(7) Lepetic used this theory to explain the collectorless flotation of dry-ground chalcopyrite ore.

Ravitz and Porter were, on the other hand, in favour of the natural hydrophobicity theory and showed that galena cleaned of its surface oxidation products floated completely in the absence of oxygen using only a frother.(8) Fuerstenau also showed that clean samples of galena, pyrite and chalcopyrite floated almost completely in the absence of oxygen, while the floatability of clean sphalerite was considerably less.(9)

During the course of investigating the effect of surface oxidation of sulphide minerals on xanthate flotation, it was found that some minerals, notably chalcopyrite, floated well in a microflotation cell without collector of frother, once the surface was cleaned of oxidation products. Consequently, it was decided to investigate whether or not chalcopyrite in natural ore samples could be made floatable by using sodium sulphide as a cleaning agent to remove the oxidation products. A literature search indicated that this reagent had not specifically been used to promote the collectorless flotation of sulphides although it was frequently used to clean sulphide mineral surfaces for the purposes of fundamental research.

DETAILS OF INVESTIGATION

Copper Ore from a Prospect in B.C.

The initial work was done on a copper ore from a prospect in B.C. The ore samples had been left over from a previous investigation carried out in 1973(10) and had been stored at the minus 10 mesh size without any precautions taken to prevent possible oxidation of the sulphide minerals. The ore contains 2.5% chalcopyrite and 20% pyrite with quartz, mica and calcite as the major gangue minerals.

A series of flotation tests was carried out with varying amounts of technical-grade sodium sulphide added to the grind. A relatively coarse grind of 96% minus 100 mesh was employed in each test. After the addition of 0.88 lb per ton Dowfroth 250 the ground sample was conditioned for 10 minutes and a copper rougher concentrate floated for 4 minutes. This was followed immediately by one stage of cleaning without reagents. The cleaner froth was skimmed until it was barren which required a skimming time of 3 to 4 minutes.

The pH measured at the end of the conditioning period varied from 8.3 for an addition of 0.4 lb per ton Na_2S to 10.5 for a Na_2S addition of 3.8 lb per ton. It should be noted that there was considerable drop in pH at the end of rougher flotation which is an indication that oxidation was taking place.

In Figure 1 the rougher flotation results obtained for the copper ore are plotted against the amount of Na_2S added. The best grade of copper rougher concentrate - 8.8.% copper was obtained using a Na_2S addition of 3.0 lb per ton. This was accompanied by a recovery of 76.8%. Higher additions of Na_2S resulted in a rapid drop in concentrate grade because of the flotation of excessive amounts of pyrite.

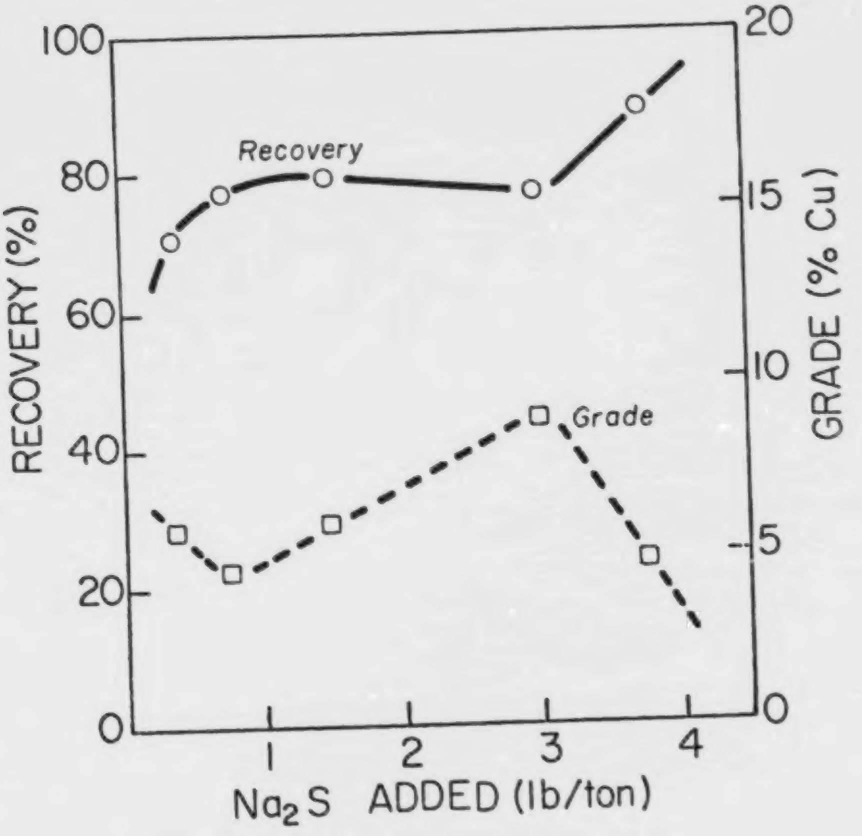


FIG. 1 EFFECT OF SODIUM SULPHIDE ON THE COLLECTORLESS FLOTATION OF COPPER ORE

Selectivity during collectorless flotation was improved considerably by lowering the pH to 6.0 with HCl after sulphidization. Table 1 gives the results of a test conducted at the lower pH. These results are superior to those reported in the previous investigation using conventional flotation techniques. At a similar grind and using potassium anyl xanthate as collector a 20.0% copper concentrate was produced with a recovery of only 38.7%.

TABLE I: Collectorless Flotation of Copper Ore at pH 6.0 by using 3.76 lb/ton of Na_2S.

Products	Weight (%)	Assay (% Cu)	Distribution (%)
Concentrate	2.58	21.35	68.08
Cleaner Tail	2.39	3.65	10.78
Tail	95.03	0.18	21.15
Feed	100.00	0.81	100.00

Texasgulf Ore

Encouraged by the results obtained for the copper ore the work was extended to the selective flotation of chalycopyrite and sphalerite from Texasgulf ore.

Figure 2 gives the flowsheet and reagent balance for the test (Test TA-7) which gave the best results using the collectorless flotation technique on the Texasgulf copper-zinc ore.

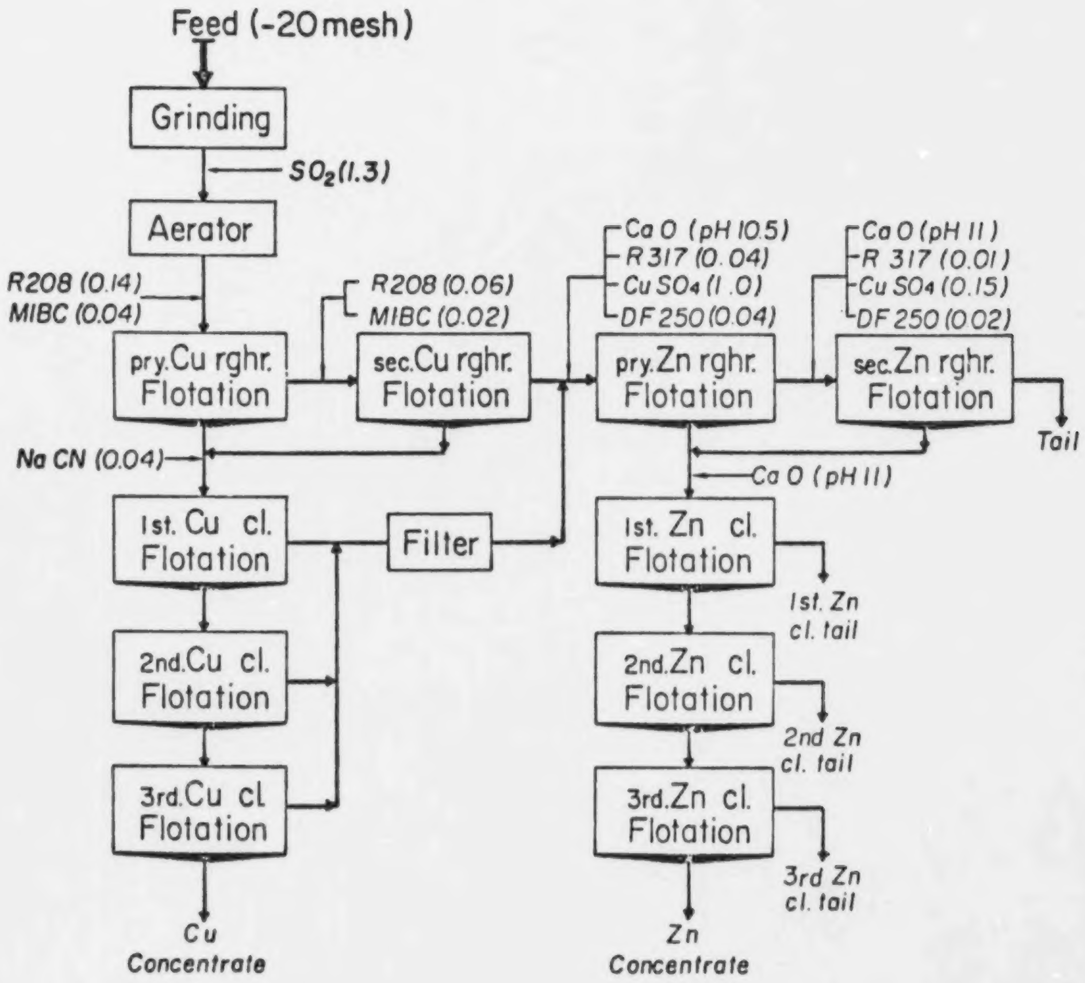

Feed (-20mesh)
Grinding
SO_2(1.3)
Aerator
R208 (0.14)
MIBC (0.04)
pry.Cu rghr. Flotation
R208 (0.06)
MIBC (0.02)
sec.Cu rghr. Flotation
Na CN (0.04)
Ist. Cu cl. Flotation
Filter
2nd.Cu cl. Flotation
3rd.Cu cl. Flotation
Cu Concentrate
Ca O (pH 10.5)
R 317 (0.04)
Cu SO4 (1.0)
DF 250 (0.04)
pry.Zn rghr. Flotation
Ca O (pH 11)
R 317 (0.01)
Cu SO4 (0.15)
DF 250 (0.02)
sec.Zn rghr. Flotation
Tail
Ca O (pH 11)
Ist. Zn cl. Flotation
Ist. Zn cl. tail
2nd.Zn cl. Flotation
2nd Zn cl. tail
3rd.Zn cl. Flotation
3rd Zn cl. tail
Zn Concentrate

Fig. 2. Laboratorv flowsheet and reagent balance (lb/ton) of test TA-9.

TABLE 2. Results of Test TA-7 and TA-8

Test No. and Remarks	Products	Weight (%)	Assays (%)			Distribution (%)		
			Cu	Pb	Zn	Cu	Pb	Zn
Test TA-7 Collectorless flotation after aerative conditioning with SO_2.	Cu conc	8.84	23.71	0.35	3.54	78.45	29.15	4.09
	Zn conc	9.38	1.94	0.25	52.50	6.81	22.17	77.66
	3rd Zn cl tail	1.06	3.52	0.53	14.72	1.40	5.28	2.46
	2nd Zn cl tail	4.21	2.03	0.21	7.82	3.20	8.30	5.19
	1st Zn cl tail	13.26	1.28	0.09	2.71	6.35	11.23	5.67
	Tail	63.25	0.16	0.04	0.41	3.79	23.87	4.93
	Feed	100.00	2.67	0.11	6.34	100.00	100.00	100.00
Test TA-8 Collectorless flotation without aerative conditioning with SO_2.	Cu conc	8.91	23.24	0.38	6.53	81.80	4.97	8.91
	Zn conc	13.78	1.34	0.19	37.33	7.29	3.84	78.77
	3rd Zn cl tail	1.59	3.13	0.27	12.31	1.97	63.07	3.02
	2nd Zn cl tail	1.02	1.64	0.16	5.49	0.66	23.83	0.85
	1st Zn cl tail	9.83	1.01	0.10	2.97	3.92	1.44	4.47
	Tail	64.86	0.17	0.03	0.40	4.36	2.86	3.97
	Feed	100.00	2.53	0.10	6.53	100.00	100.00	100.00

As in conventional flotation the ground ore was aerated with SO_2 prior to copper flotation. All reagents including Na_2S and $CuSO_4$ were stage fed. Higher molecular weight Dowfroth 1400 was used in place of the Dowfroth 250 employed for the copper ore. Also small amounts of kerosene were added. It was found that the kerosene speeded up the zinc flotation rate and improved selectivity. The three copper cleaner tails were combined, filtered and added to zinc rougher flotation. A small amount of $KMnO_4$ was added to the first zinc cleaner for pyrite depression.

In Table 2 the results of this test are compared to a parallel test in which the aeration step with SO_2 was omitted. As can be seen the main effect of the aeration step with SO_2 was to greatly increase zinc concentrate grade without adversely affecting recovery; 52.5% zinc, 77.7% recovery for aerative conditioning with SO_2 as against 37.33% zinc with 78.9% recovery when the aeration step was omitted. Also aerative conditioning with SO_2 cut the zinc loss in the copper concentrate in half, from 8.9% to 4.1%.

Conventional Test Procedure

To compare with collectorless flotation a test was conducted using conventional flotation procedure with thiol-type collectors. The reagent balance used at the Texasgulf concentrator was followed as closely as possible. Test procedure is given in Figure 3. Note that the flowsheet is identical to that employed in the collectorless flotation tests. The thiol-type collectors employed were R 208 (a dithiophosphate) for copper flotation

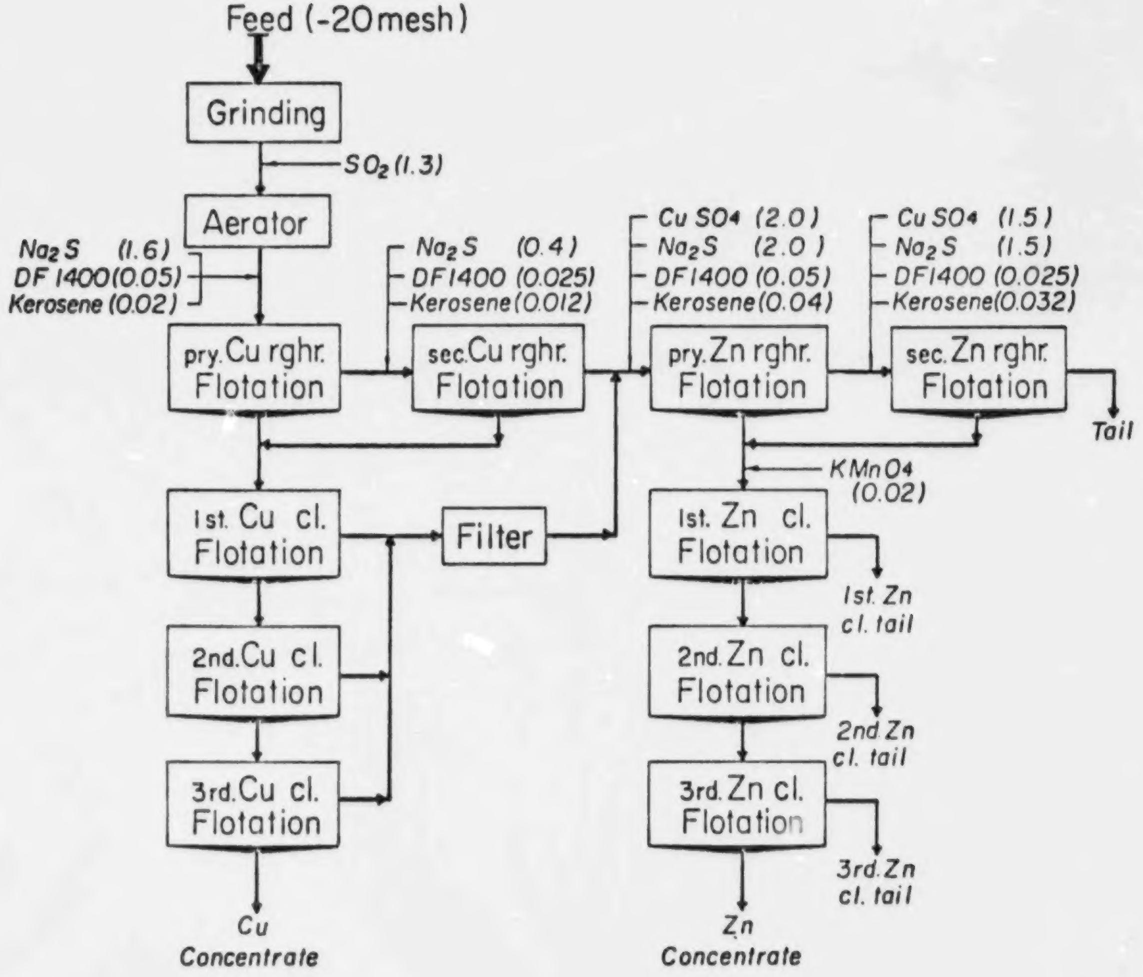

Feed (-20mesh)
Grinding
SO2 (1.3)
Aerator
Na2S (1.6)
DF 1400 (0.05)
Kerosene (0.02)
pry. Cu rghr. Flotation
Na2S (0.4)
DF1400 (0.025)
Kerosene (0.012)
sec. Cu rghr. Flotation
Cu SO4 (2.0)
Na2S (2.0)
DF1400 (0.05)
Kerosene (0.04)
pry. Zn rghr. Flotation
Cu SO4 (1.5)
Na2S (1.5)
DF1400 (0.025)
Kerosene (0.032)
sec. Zn rghr. Flotation
Tail
1st. Cu cl. Flotation
Filter
KMnO4 (0.02)
1st. Zn cl. Flotation
1st. Zn cl. tail
2nd. Cu cl. Flotation
2nd. Zn cl. Flotation
2nd Zn cl. tail
3rd. Cu cl. Flotation
3rd. Zn cl. Flotation
3rd. Zn cl. tail
Cu Concentrate
Zn Concentrate

Fig. 3. Laboratory flowsheet and reagent balance (lb/ton) of test TA-7.

TABLE 3: Results of Test TA-9

Test No. and Remarks	Products	Weight (%)	Assays (%)			Distribution (%)		
			Cu	Pb	Zn	Cu	Pb	Zn
Test TA-9 Flotation using R 308 and R 317 after aerative conditioning with SO_2.	Cu conc	8.54	25.48	0.47	4.68	83.70	39.88	6.19
	Zn conc	9.73	2.82	0.09	52.96	10.55	8.73	79.78
	3rd Zn cl tail	0.79	1.89	0.28	11.80	1.10	3.07	3.43
	2nd Zn cl tail	1.39	3.62	0.39	28.05	1.01	3.87	2.54
	1st Zn cl tail	5.56	0.64	0.14	3.24	1.37	7.74	2.79
	Tail	73.99	0.08	0.05	0.46	2.27	36.71	5.27
	Feed	100.00	2.60	0.10	6.46	100.00	100.00	100.00

and R 317 (sodium isobutyl xanthate) for zinc flotation. Table 3 gives the results of conventional flotation using collectors. The copper concentrate grade and recovery were higher than those obtained for collectorless flotation, 25.5% copper versus 23.7% copper, 83.7% recovery as against 78.5% for collectorless flotation but zinc loss in the copper concentrate was higher, 6.2% versus 4.1% for collectorless flotation. The zinc results were comparable to those obtained for collectorless flotation.

Collectorless Flotation without Na_2S

To demonstrate that Na_2S was essential for collectorless flotation one test was carried out on the Texasgulf copper-zinc ore in which the addition of this cleaning agent was omitted except in the zinc scavenger. Table 4 gives the results of this test. After one stage of cleaning a copper concentrate grading 16.7% copper was obtained with 71.8% recovery as compared to the 23.7% grade and 78.5% recovery obtained with Na_2S. It should be pointed out, however, that about 5 times as much frother was added in the test without Na_2S and the chalcopyrite flotated so slowly that it required 45 minutes of skimming in the copper rougher.

In the case of zinc flotation, a low-grade zinc concentrate was produced (36.4% zinc) with a recovery of only 28.5%. Upon the addition of Na_2S to the zinc scavenger and further activation with copper sulphate an additional 39% of the zinc was floated in a zinc concentrate grading 39.9% zinc.

TABLE 4: Results of Test TA-10

Test No. and Remarks	Products	Weight (%)	Assays (%)			Distribution (%)		
			Cu	Pb	Zn	Cu	Pb	Zn
Test TA-10 Na_2S was used only for Zn scavr. flotation.	Cn conc	11.04	16.68	0.35	7.16	71.76	37.59	12.35
	Zn conc	5.01	5.26	0.37	36.35	10.27	18.01	28.46
	Zn cl tail	5.76	1.55	0.12	6.39	3.48	6.27	5.75
	Zn sc cl conc	6.22	1.05	0.12	39.87	2.54	7.30	38.75
	Zn sc cl tail	4.91	1.19	0.09	5.49	2.28	4.28	4.21
	Tail	67.06	0.37	0.04	1.00	9.67	26.10	10.48
	Feed	100.00	2.57	0.10	6.40	100.00	100.00	100.00

DISCUSSION

It has been established in this investigation that both chalcopyrite and sphalerite can be floated selectively from other sulphides without using a thiol-type collector if the ore is treated with a cleaning agent such as Na_2S to remove the surface oxidation products. Although it was not the purpose of the investigation to prove whether or not sulphide minerals are intrinsically hydrophobic, the results obtained appear to support this view. It may be fruitful, therefore, to discuss some of the possible causes for the apparent hydrophobicity of sulphide minerals.

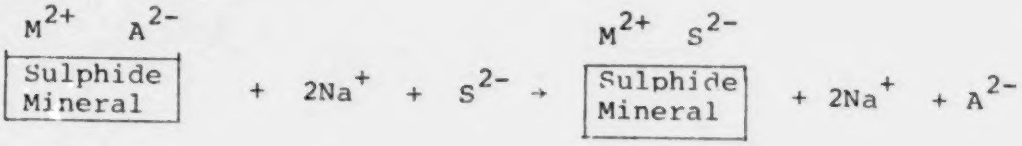
First, one must consider the possible contamination of ore samples during grinding and flotation or by the impurities in the reagents used. It is difficult, however, to conceive that contamination of such origins could account for the reproducible results of the flotation tests obtained for large quantities of ore samples.

Second, elemental sulphur, a well-recognized hydrophobic entity may be deposited on the surface by the oxidation of sulphide minerals themselves or by the oxidation of sulphide ions introduced as Na_2S. This is unlikely however since both chalcopyrite and sphalerite float well at an alkaline pH which is not thermodynamically favourable for elemental sulphur formation.

Finally, the frothers used may have collecting power. Heyes and Trahar who used Polypropylene glycol 400 in their collectorless tests with chalcopyrite were skeptical, however, of its collecting power because it has no functional groups(11). Although a frother is not likely to induce hydrophobicity, it could absorb through van der Waals forces on hydrophobic surfaces with the strength of adsorption depending on the molecular weight or length of the hydrocarbon chain. In this limited sense the forthers used in collectorless flotation could be regarded as having collecting power.

Cleaning Action of Na_2S

To explain the cleaning action of Na_2S this ion exchange mechanism is proposed:



$$\overset{M^{2+}\quad A^{2-}}{\boxed{\begin{matrix}\text{Sulphide}\\\text{Mineral}\end{matrix}}} + 2Na^+ + S^{2-} \rightarrow \overset{M^{2+}\quad S^{2-}}{\boxed{\begin{matrix}\text{Sulphide}\\\text{Mineral}\end{matrix}}} + 2Na^+ + A^{2-}$$

The M^{2+} represents the metal ion of the sulphide while the A^{2-} represents the anionic oxidation products such as sulphates, various thiosalts, oxides and hydroxides. The equation will proceed from left to right because most of the oxidation products are much more soluble than the substrate sulphide mineral. The sulphide mineral cleaned of its hydrophylic oxidation products is thus restored to its natural hydrophobic state.

Selective Flotation of Chalcopyrite from other Sulphides.

In some previous work on the effect of oxidation on sulphide flotation it was found that both chalcopyrite and pyrite when cleaned by freshly-prepared ammonium sulphide exhibited 100% floatability in a microflotation cell without the aid of a frother. The question arises therefore, as to why it was possible in this investigation to float chalcopyrite selectively from ore samples which contained 20 to 35% pyrite. The answer to this question is related to the solubility of the common sulphides.

In Figure 4 the solubility of various sulphides in water and in the absence of oxygen are plotted as a function as pH. As can be seen covellite and chalcopyrite are the least soluble of the sulphides. Therefore, if a mixture of common sulphide minerals is treated with a Na_2S solution of relatively low concentration, the copper minerals will be sulphidized first and float in preference to the other sulphides assuming that all sulphides are oxidized to the same extent. The copper-activated sphalerite will of course behave like covellite in this respect.

Another possible explanation for the selective flotation of chalcopyrite is that it does not oxidize to the same extent as other sulphides in the ore. If this is the case, its sulphidization would be completed ahead of the other sulphide minerals and it would float first.

CONCLUSIONS

It can be concluded that collectorless flotation using Na_2S as a cleaning agent appears to be a viable alternative to conventional sulphide flotation with thiol-type collectors. Although Na_2S is cheaper (about 12¢ per lb as against 55¢ per lb for collectors) the amounts used in this investigation were such that the total cost was greater than the cost for collectors. However, no attempt was made to optimize Na_2S additions. Theore-

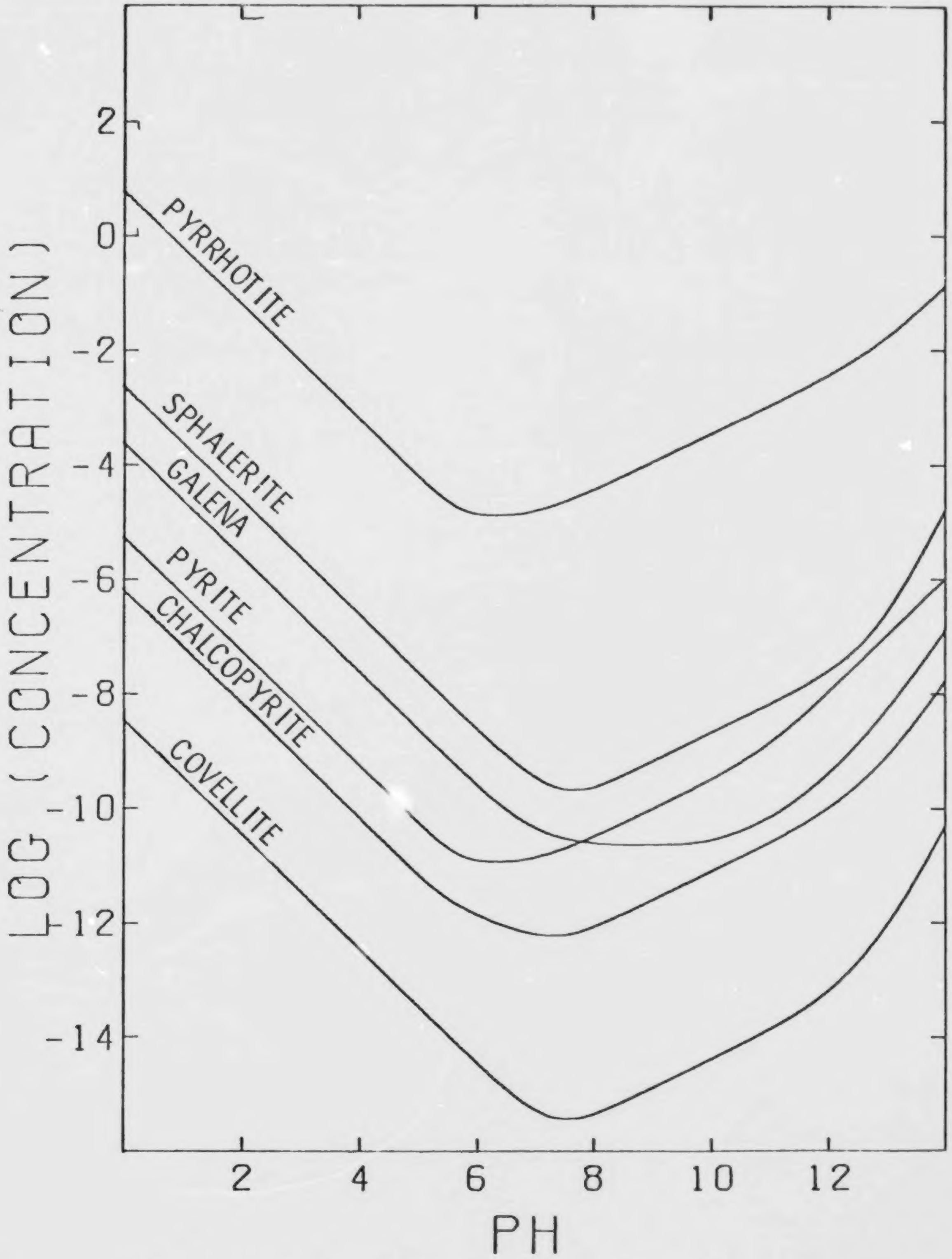


FIGURE 4: Total solubility of metal sulphides in the absence of oxygen as a function of pH (closed systems)

tically, sulphide ion comsumption should not be greater than collector consumption. If this is so, collectorless flotation would be competitive in cost with conventional flotation using collectors.

REFERENCES

1. Boyce, J.H. Venter, W.J. and Adam, J., World Symposium on the Mining and Metallurgy of Lead and Zinc, Vol. 1, p. 542, 1970.
2. Lepetic, V.M., C.I.M. Bulletin, 71, June 1974.
3. Sulman, H.L. I.M.M. Bulletin, No. 311, August 1930.
4. Gaudin, A.M. "Flotation", McGraw-Hill Book Co. Inc, New York, p. 38, 1932.
5. Taggart, A.F. del Guidice, G.R.M. and Ziel, D.A. Transactions, AIME 112, p. 348, 1934.
6. Wark, I.W. "Principles of Flotation", 1st edition, p. 127, 1938.
7. Plaksin, I.N. Dokl. Akad, Nauk SSSR, 16. No.1, 91, 1949.
8. Ravitz , S.F. and Porter, R.R., AIME Tech. Pub. No. 513, 1933.
9. Fuerstenau, M.C. "Future Trends in Mineral Processing Research" special lecture, McGill University, Montreal, November 1975.
10. Raicevic, D, and Bruce, R.W. Mines Branch Investigation Report, IR 73-50, May 1973.
11. Heyes. G.W. and Trahar, W.J., "Floatability of Chalcopyrite", unpublished report, C.S.I.R.O., Clayton, Victoria, Australia

Discussion of Paper by R. Yoon & A. Stemerowicz

G. Holmes: You say that this is a viable alternative! Are there any special tricks with it ?

Reply: It might be especially suitable for oxidized ores. You might also add small amounts of collector.

Paper No. 11

ARSENIC IN PYROMETALLURGICAL OPERATIONS

by

John N. Cram

ABSTRACT:-

The comments given are to help allay generally accepted ideas that:- the handling of arsenic as arsenic-trioxide (As_2O_3) in and from off-gas pyrometallurgical operation fumes is difficult, hazardous and expensive. It need not be any of these things if the basic and rather unique chemical and physical properties are understood and taken care and advantage of through present day technologies. These particular characteristics are:-

(1) As_2O_3 sublimes and re-condenses, at a relatively low temperature;

(2) As_2O_3 is, for practical in-plant purposes, insoluble in water at ambient temperatures and quite soluble in water just below the boiling point;

(3) residual suspended and dissolved As_2O_3 can be removed from water by complete dissolution with very little sulfuric acid, followed by the addition of small amounts of ferrous sulfate and calcium chlorate, with oxidation to the ferric and arsenite state and precipitation of an insoluble ferric calcium arsenate, thereby lowering the As_2O_3 content of solutions to acceptable levels. Even lower levels of non-toxic forms of arsenic solids and residual As_2O_3 in the water solutions can be attained by also using phosphate, and lime, along with the above.

A background re past arsenic processing methods, present day operations, plus predictions re future trends, and hazards, are outlined.

ARSENIC - PYROMETALLURGICAL OPERATIONS

GENERAL BACKGROUND:-

Most arsenic containing precious metal and many other non-ferrous concentrates have been in the past and still are usually subjected to roasting during their processing to eliminate most of the sulfide and/or arsenic content. This facilitates and improves metal extraction or recovery and thereby the overall economics. The roasting is usually two stage - first, with a reducing atmosphere followed by an oxidizing stage to emit fumes with sulfur-oxides and arsenic-trioxide (As_2O_3) as gases, plus the carried-over fine solid particles of gangue, concentrates and perhaps even metal particles.

The long established procedures for processing straight arsenic-bearing and/or arseno-pyrite concentrates was, until relatively recent years, and still is in some operations, to cool the flue gas fumes and thereby condense the contained As_2O_3 gases, collect all the solids in the cold fabric dust collectors. The collected fine dust was termed 'crude arsenic' and usually ran from 50 to 85% As_2O_3 plus fine particles of gangue and concentrates. This crude arsenic could not normally be sold as collected, as the As_2O_3 content was too low for most major industrial uses. Relatively small quantities were sold and used to produce herbicides, insecticides and products such as rat poison. These products are no longer allowed to be produced in Canada.

The crude arsenic could not be sold as it contained values and gangue, and therefore it had to be disposed of by dumping. This was costly if the ore-concentrates came from precious metal mines and the dollar value of the contained values was considered. Nevertheless, large tonnages of it were dumped, and some probably still is being dumped in out-of-the-way places, underground, up in mountains and in concrete or other type storage bins. It has been rumoured that one European producer took As_2O_3 out to sea in concrete bins and dumped it. In any case, some of these past disposal methods have or may well come back to haunt us - because of expected or imagined potential long term environmental hazards. There is talk that some operators will have to, or will of their own accord, bring up some underground stored crude arsenic for re-processing, upgrading or even to be sold for special uses.

Those mine operators and/or extractive metallurgical plants that did, in days past, 'refine' their 'crude arsenic' and obtain a high grade saleable As_2O_3, as well as recover the values, usually did so by the traditional procedure of pyro (fire) refining the crude arsenic with a soda-ash flux in reverbatory furnaces at over-fire temperatures of around 2,200°F. The As_2O_3 content re-sublimed, passed off through flues to a series of chambers 'kitchens', and as it cooled, condensed out and settled as a high grade As_2O_3 product in the kitchens (except for the first few and last few where some carried-over gangue and concentrate solids settled out. The line of kitchens was periodically shut down, cleaned out, (often by manual labour), the material packed into drums and shipped to market. The slag from the reverbatory furnace was tapped periodically and returned to the furnaces and/or roasters for recovery of the values.

The price received for the 99% As_2O_3 for many years (in the late 1940s and well into the 1950s) often did not cover the cost of the drums and freight to market; during World War I, the price rose considerably (reportedly to 60¢/lb.) as arsenic was used in the shell casings (steel alloys) to make them brittle, thus, the word 'shrapnel'. Again during the more recent Vietnam War, the price rose as As_2O_3 was used to produce defoliants for use on the jungle-type vegetation of Vietnam. The writer, when at Deloro, saw sale prices from as low as 2¢/lb., in the late 1950s to 32-34¢/lb. in the mid and late 1960s. Present prices run 20 - 25¢/lb. f.o.b. producer plant.

Deloro Smelting & Refining (owned by the O'Brien's of Ottawa) processed the arsenical silver-cobalt ores from Cobalt at Deloro. Shortly after the discovery of silver at Cobalt, Ont., Senator M.J. O'Brien along with others purchased the then closed down arseno pyrite gold ore processing plant at Deloro. The main reason for the purchase was that the plant had roasting facilities to handle the arsenical concentrates from Cobalt. It ran for about 60 years and up to closure collected their crude arsenic and refined it in a reverb type arsenic refinery, complete with kitchens, etc.

Similar arsenic refineries were used in the past by several large U.S. and European metallurgical plants.

After Deloro Smelting & Refining Co. closed in 1960 and operations were initiated in 1961 at the closed down Cobalt Chemicals plant near Cobalt, Ontario, as The Cobalt Refinery, the only feed to prove out the circuits was some 900 tons of crude arsenic left by the previous operators. This was leached and/or treated with caustic soda to obtain upgraded silver bearing residues. These in turn were used as the first feeds to the roaster and smelting furnace. The bottoms were fire refined to a base bullion and the circuits and the process were thus proven out. Some of the crude arsenic was subjected to direct roasting, the As_2O_3 condensed under close control and the fumes collected in a cold baghouse, packed and sold as high-grade As_2O_3. As a result of these procedures, some silver-cobalt concentrates were received and fine silver was produced, but in so doing more crude arsenic was collected off the roaster. There was no pyro-arsenic refinery. A small reverbatory furnace was built and used for several years. Instead of the traditional cooling chambers, the 'roaster' fabric cold dust-collector was used. This, at best, was a poor arrangement as roasting and arsenic refining could not both be run at the same time. Labour turnover was very high, mainly because of arsenic skin burns. It was then decided to attempt to utilize the low temperature sublimation property of As_2O_3 to collect the solids from the flue-gas fumes separately on their own, then cool the gases and hopefully re-condense the clean As_2O_3 and catch it in a cold baghouse. Wheelabrator Corporation of Canada was approached and agreed to supply free a test unit with 12 specially treated fibreglass fabric dust tubes. This was placed in a by-pass flue off our main roaster flue with the outlet to a separate cold fabric dust-collector. The test run results were good. A production unit was ordered and installed between the roaster and the previously installed cold fabric dust-collector. It was set up with a temperature controlled damper in the roaster flue ahead of the new hot baghouse and set to control temperature in the hot baghouse at 500 - 525°F. by bleed-in ambient air. This was the maximum temperature that the fibreglass fabric then available would stand up to on a continuous basis. The system provided an upgraded revert concentrate relatively low in arsenic, and a 99% grade of As_2O_3 in the cold collector was collected separately. The system worked well providing the maintenance was checked regularly and if holed bags in the hot collector were replaced. The main problems encountered were that the glass fibres, typical of glass, tended to break from the frequent flexing, particularly near the bottom collar just above the floor plate. Further, the 500 - 520°F. temperature for the hot collector was not really high enough to ensure that the As_2O_3 remained gaseous and at that time, 1965-66, we were not able to obtain a fibre that would stand up to continued high temperatures, although Manhattan Raybestos had a re-inforced asbestos fabric under development. As a point of interest, nylon type bags failed within a few days in the cold baghouse. This was due to acid hydrolysis from SO_2 and/or SO_3 carried over from the roaster. Specially treated terylene bags gave much better performance. In any case, the reverbatory furnace arsenic refinery was closed down; operating costs for labour, W.C.B., fuel and repairs were lowered, and reduced labour turnover improved morale; availability of roasting facilities increased. The crude arsenic manual collection, movement and feeding to the hot furnace hoppers was eliminated, as also was tapping and handling of the hot arsenic refinery furnace slag and crushing for return to smelter furnace. The upgraded fine solids from the hot collector could be returned as collected directly to the furnaces. To the writer's knowledge, this was the first of such operations installed and operated. Since then, Dickenson Mines has installed a similar system at their Balmertown mill. It is much more sophisticated and has better controls. 'Hot' fabrics for the 'hot' collector are available from Germany and it is believed are also from The U.S.A. These bags will stand temperatures up to 650°F. or higher but the operating life still leaves much to be desired. They are expensive, but the secret of good, clean separation and a high grade As_2O_3 product is close control and maintenance including replacement of bags, etc. If these precautions are followed, the As_2O_3 collected will be saleable at regular market prices and this, as already stated, means low labour costs, safe handling and no dumping or arsenic refining problems. Operating costs can be reduced, the As_2O_3 health hazard, both immediate and long term, can be eliminated, particularly if a completely closed pneumatic collection, movement and packaging system is installed. Capital costs for such installations are not exorbitant.

GENERAL CHARACTERISTICS - Arsenic-Trioxide (As_2O_3

As_2O_3 has two (2), not too often found, basic chemical and physical properties which, if taken advantage of, greatly facilitate the handling, collection and upgrading of the arsenic content of ores and/or concentrates during pyrometallurgical operations.

These properties are:-

(1) As_2O_3 sublimes at relatively low temperatures (218°C - 425°F) from the solid state direct to the gaseous state and vise-versa.

(2) As_2O_3 is for practical operating purposes insoluble in water at normal ambient temperatures but becomes quite soluble, up to 100 grams/litre, at or near boiling (95°C - 200°F).

Mineral Processing & Arsenic:- Today in Canada.

Roasting of arseno-pyrite or arsenide ores and/or concentrates is usually carried out in two stages (whether in a single, double roaster or in two separate roasters). Two separate roasters lend themselves to better control. The first stage is under a reducing atmosphere, followed by an oxidizing atmosphere stage. The result is, most of the arsenic content goes off with the flue-gas fumes as As_2O_3 in the gaseous state. Nowadays, much of the carryover of fine solids is caught in hot cyclones, and the remainder is collected in a 'hot' electrostatic precipitator or a 'hot' fabric dust-collector.

Most of the dusts with metal values are collected in the hot precipitator, and/or almost all is caught in the dust tubes if a 'hot' fabric dust-collector is used. This assumes either or both types of equipments are functioning efficiently. The As_2O_3 as a gas passes through either type of hot collector, is then cooled by either bleed-in ambient air or with air-cooled heat exchangers. The As_2O_3 then condenses back to solid As_2O_3 particles (crystals) and is collected in a cold baghouse at 120^o - 150^oC (250^o - 300^oF) with the 'arsenic clean' flue gases passing through to the stack. In either case, the temperature of the cold fabric dust-collector must be kept as much below the As_2O_3 sublimation temperature as practical but above the dew-point. The dew-point can vary dependent on several factors and in particular the sulfur content of the feed ore or concentrates, the oxidation state of the sulfur as oxides and the resultant SO_2 and SO_3 content and ratio in the 'cold' flue gases. It would seem that higher sulfur oxides content, and especially SO_3, can necessitate higher temperatures in the 'cold' baghouse. This in turn can result in higher As_2O_3 vapour pressures and carry-through of some extra fine As_2O_3 particles to the stack. The overall result may well be that, with the hot precipitator and cold fabric dust-collector, as opposed to the hot and cold fabric dust collectors, the grade of As_2O_3 from the hot precipitator system will generally be of lower purity than from the hot and cold fabric dust-collector system, but the final gas emissions to the stack may be slightly higher in As_2O_3 from the hot and cold baghouse system. This is of course with the assumption that both types of systems are well maintained and operating efficiently.

Both systems of hot collection of values and cold collection of high quality (99% As_2O_3) have advantages and disadvantages. Both the hot precipitators and/or hot baghouses must be well insulated to maintain the temperature throughout (up to 345^oC - 650^oF or slightly higher).

If the 'hot' precipitator is not adequately insulated, some As_2O_3 will condense on the walls and eventually on the electrodes with shortings and a major loss of efficiency. The As_2O_3, once crystallized, will precipitate out as does the gangue and metal values and as a result will not as efficiently collect the fine solid particles of gangue and/or concentrates (values) as the capacity is used up collecting re-condensed As_2O_3. The same, of course, can happen in hot fabric dust-collectors. The second disadvantage of the hot electrostatic precipitator is that they do not precipitate or collect all the solid particles. Some of the gangue and concentrates pass through to the cold fabric collector so that the As_2O_3 collected is 'off-grade' and not acceptable by many industrial users.

The hot fabric dust-collector functions like a wall. If the bags (dust tubes) are not holed, literally all the solid values are collected in it and the As_2O_3 is collected in the 'cold' baghouse and will be of 99% quality.

The capital cost of the hot precipitator system is higher and its As_2O_3 disposal problems by marketing and/or waste disposal are complicated and expensive with low return, if any, on capital and/or operating costs. The hot fabric dust-collector should have a lower capital cost. Disposal is simplified as the produced (high grade As_2O_3) can be sold to industrial users at a fair price. The dust tubes are expensive and do result in high maintenance costs; nevertheless, there are less disposal problems as the product can be sold to industrial users and should create a fair cash credit to the overall cost picture.

RE-PROCESSING OLD STOCKPILES AND/OR CURRENTLY PRODUCED & COLLECTED CRUDE ARSENIC:-

As stated, it is expected and known to some extent that there are in Canada and in The U.S.A., sizeable tonnages of crude arsenic containing some values than will have and/or may soon have to have something done with them for two main reasons; (1) increased values of metal contents particularly gold and silver, and (2) for environmental reasons. Environment Canada, E.P.A. in The U.S.A., Provincial Ministries of The Environment, all are thinking of and concerned about the so-called dumps of crude arsenic from past operations - whether buried, out of sight or otherwise. The public is being stirred into concern also and without adequate knowledge by the public, media and others, a storm of concern has been and is brewing. There are also indications that permission to store and/or dump current production of crude arsenic will be withdrawn in the foreseeable future. This means that sooner or later some action will more than likely be forced or taken voluntarily to clean up such public and governmental concerns.

It would seem that the simplest way to process old crude arsenic stockpiles or current production is to take advantage of the great differential in solubility of As_2O_3 in cool and hot water. The crude arsenic can be leached in tanks with agitated hot water; the As_2O_3 will be dissolved, and the hot slurry then filtered to give two products:-

(1) the undissolved residue, low in arsenic, which is the upgraded gangue and concentrates with mineral values, amenable to metal extraction;

(2) the As_2O_3 saturated hot-leach solution which when cooled becomes supersaturated with the As_2O_3 crystallizing out. This can then be filtered off, the crystals washed, centrifuged, dried and packed as 99% grade As_2O_3 and sold to industrial users in The U.S.A. and other parts of the world.

A facility such as this is being organized, and hopefully will be set up at the old Cobalt Refinery plant. It is also understood that

Giant Yellowknife is seriously considering similar action.

This type of operation should be straight forward, produce saleable products and clean up existing and potential environmental problems.

HAZARDS:

Some general comments about the potential, real and over-imagined hazards of working with and/or in contact with arsenic-trioxide would seem to be in order. Arsenic metal, arsenic pentoxide and arsenates are not toxic. Arsenic-trioxide is toxic even in quite minute quantities if ingested, although persons working with it or ingesting it in small amounts can and do build up an increased tolerance. Some persons are, as with most toxic materials, more susceptible and/or allergic than others; generally, persons with light or reddish complexioned skins are more allergic or sensitive to burns.

There are three main working hazards:-

(1) Breathing the dust: there are claims that it can and does cause lung, stomach or general digestive system cancers. There were Ontario W.C.B. awards for lung cancer and emphysema, particularly for ex-employees of Deloro Smelting & Refining during the 1950s at Deloro, Ontario (now closed). Even so, there is, as understood by the writer, no conclusive evidence that lung cancer is actually caused by arsenic-trioxide. It is more than likely that continued exposure to arsenic-trioxide dusts can contribute to the onset of emphysema, especially along with regular cigarette smoking. This can be so with any dusts. There is some expert opinion that the Deloro lung cancer in the award cases attributed to arsenic-trioxide may well have been due to other non-ferrous oxides exposure, perhaps nickel-oxides. Awards have been made at other operations for lung cancer and attributed to nickel-oxides. Nonetheless, special precautions should be compulsory and strictly enforced with employees involved in the collecting, handling and packaging of arsenic-trioxide (crude or refined). They should, if

possible and practical, wear protective hoods with a positive clean air supply. This is a common practice in many non-arsenic-trioxide operations. On the other hand, closed system collecting, moving and packing, can and will to a large degree obviate the dust inhalation hazard and perhaps the elimination of the need for such hoods.

It is noted that air ventilated head hoods work well when used inside heated buildings but are almost useless in sub-zero - outside weather. The workers exhale, their breath is warm and high in moisture vapour, it condenses inside the cold hood, and fogs up the face piece and the worker cannot see. On the other hand, there are complete air-ventilated sealed suits. They certainly will offer complete protection if practical for use. They are used in many other operations - to protect from toxic gases, liquids or dusts. Some of these suits, however, are so heavy and awkward they are not very practical for a worker who has to move about and use his arms for lifting and handling. It is understood there are now complete fully sealed suits that weigh only about 10 pounds.

The 'Hazard' comments for workers have so far referred to work with As_2O_3 as a dust or fume.

(2) The problem of working with As_2O_3 solids in solution or in a wet stage presents similar possible hazards. The above comments are even more applicable re contact with As_2O_3 or arsenite solutions. This is particularly so during leaching operations and especially if the pH of the solutions is either basic or acidic. Workers in direct contact with such conditions should have complete body protection and change same frequently, unless it is a sealed ventilated suit.

All As_2O_3 workers should always have a complete shower when changing from work clothes to their street clothes. There should also be facilities including separate lockers for work and street clothes.

Another useful idea is to have a centre or rough shower cubicle where outer work suits, hoods, mitts, gloves, etc. can be given a good wash-down with real hot water to wash and/or leach off the As_2O_3. This wash water run-off should drain to a sump and be pumped into a holding or leach tank rather than into the domestic or sanitary sewers. It can then be treated or leached.

(3) Skin burns and/or rashes: from contact with arsenic-trioxide can be very unpleasant. Open cuts and/or bruises if contacted by arsenic-trioxide tend to absorb it, as they often are moist and are either acidic or basic. This action is accelerated by heat. Therefore, workmen must wear proper protective clothing and follow the proper personal hygiene procedures or they will likely end up with an arsenic rash and/or burns, even without cuts or skin abrasions. Areas around the nose, eye-lids, the groin and under-arms are very sensitive, if the worker sweats. Perspiration is acidic and tends to dissolve the arsenic-trioxide readily, forming arsenous acid. This in turn may well result in mild-to-severe burns in and around these sensitive areas. The best and almost the only realistic protection against arsenic skin burns is to practice good personal hygiene habits, particularly when workers go to relieve themselves. The arsenic-trioxide contaminated outer protective clothing should be completely removed before washing the hands. This hand washing is most important and should be done by a thorough washing with cold water - no soap, prior to contacting other parts of the skin, especially around the groin. Otherwise, if, as happens, the worker washes his hands, then opens the dirty outer and inner clothing, relieves himself, he more than likely ends up at the Doctor's office within a few days with very uncomfortable skin rashes and burns around the groin. It is similar to the saying that leading a horse to water doesn't make it drink. Instructions, with positive enforcement, re such procedures are most important. Workers should also be encouraged to take a break, even twice a shift, to remove outer and inner clothing, take a thorough cold shower, then a warm one, and then put on clean work clothes. The protective clothing should have snug ties around the ankles and outside of work boots; long sleeved mitts or gloves over the cuffs of the protective suits with elastic wrist bands will to a large degree eliminate the chance of burns to hands, wrists and arms.

The bottom flaps from the head hood should be tucked under the outer protective work suit to protect neck and head, from direct contact with arsenic trioxide dust.

Paper No. 12

LEMOINE MINES LIMITED
MINES LEMOINE LIMITEE
P.O. BOX 6600
CHIBOUGAMAU, P.Q.

MILLING OPERATIONS
AT
LEMOINE MINES LIMITED
BY
GUY LADOUCEUR - MILL SUPERINTENDENT.

LOCATION

The mine and mill are located in Lemoine township which is 20 miles directly south-east of the town of Chibougamau Quebec. Personnel live in Chibougamau and commute to the property over 33 miles of paved and gravel roads. Concentrates are trucked 13 miles to a railway siding on the line connecting Chibougamau and Quebec city.

HISTORY

Several companies explored the property before acquisition by Patino Mines (Quebec) Limited in 1973. Further exploration led to the discovery of a high grade copper-zinc orebody and the property was brought into production by the above company in 1976.

ORE

The ore contains 85% - 95% sulphides consisting of chalcopyrite, sphalerite, pyrite, pyrrhotite and minor gold and silver. Argentite, tetrahedrite, antimonial silver, native bismuth and galena are also present in very minor quantities. Associated gangue minerals are quartz, carbonates and occasionally chlorite.

The chalcopyrite exhibits a lamellar structure while the spalerite has no visible structure. The pyrite occurs as cristals and in spheroidal aggregates usually less than 1mm. in diameter but witch can occassionally be up to 5mm. in diameter. The sulphides are usually fine to medium grained but because of some remobilisation at foot-wall and hanging-wall, coarse grained sulphides do occur associated with massive white quartz.

OPERATION

The mine operates on a five day a week basis to produce between 9,000 and 10,000 tons of ore per month. Shop service is five days per week while the milling is on three shifts seven days per week. A bi-weekly downshift is provided in the mill for maintenance. The milling rate averages 350 tons per day. Separate copper and zinc concentrates are produced. The mill flowshett is illustrated in fig. 1.

CRUSHING

All crushing is done on surface as there are no underground facilities. Primary crushing to a nominal minus 3" size is by a 30" by 42" pioneer jaw crusher. This crusher is fed by short inclined conveyor belt which in turn is fed by a 4' by 18' telsmith vibrating grizzly below a coarse ore bin adjacent to the mine headframe. The vibrating grizzly has the grizzly section blocked off. Ample space is provided beneath the telsmith for front-end loader clean-up of any slime run off or spillage. The primary crusher product is conveyed to a 5' by 12' symons rod-deck screen with 3/8" by 6" openings. From the screen the undersize is conveyed to the fine ore bin while the oversize is conveyed to a second screen, a 3' by 8' rod deck. The second screen has 1" by 6" openings to separate the screened oversize into feed for a 3' standard symons cone crusher and feed for a 3' symons short head crusher. The crushed product from both crushers is collected on the same belt and returned to a 5'X12' screen thus closing the crushing circuit. The fine product is conveyed to a 1,000 ton mill ore bin.

GRINDING

Fine ore is withdrawn from the storage bin through a slot onto a speed regulated belt which in turn discharges onto the ball mill feed belt. The ore is weighed over a Merrick weightometer on the feed belt. Primary grinding of the ore is by a 9' X 8' Marcy ball mill driven by a 450 H.P. motor. The ball mill discharge is pumped by 5" X 5" SRL pump through a 15" krebs cyclone fitted with a 4" Vortex finder and 3" apex valve with the underflow returning to the ball mill and the overflow at 80% minus 200 mesh going to flotation. The ball mill circulation load is 400%. Primary grinding media is 3" forged steel balls and the mill is rubber lined. Secondary grinding of copper scavenger concentrates, copper rougher cleaner tails and copper cleaner scavenger concentrates is done in a 5' X 10' Marcy ball mill driven by a 75 H.P. motor. The mill is in closed circuit with a 10" krebs cyclone fitted with a 3" Vortex finder and 2 1/4" apex. The circulating load is 150% and cyclone overflow at 97% minus 200 mesh is sent to scavenger cleaning. The mill is rubber lined and grinding media is 1" forged steel balls.

FLOTATION

Selective flotation of chalcopyrite, after 10 minutes of conditioning the pulp from the cyclone overflow with reagents, is performed in a bank of six #24 Denver D.R. cells (Roughers) followed by a similar bank of eight cells (scavengers). The copper rougher cells are arranged such that concentrates from one or two cells may be taken directly to product at 25% copper grade or better while the product from the remaining four or more cells at approximately 15% grade of rougher concentrate is directed to copper rougher cleaning.

The copper rougher concentrates are cleaned in a bank of six #24 Denver sub A type cells to a grade of 25% copper concentrates to product. This bank of cells is arranged four and two to give two stages of cleaning. There is no recycling of concentrates in the above copper circuits except the second stage copper rougher cleaner tails through the first stage.

Copper scavenger concentrates and copper rougher cleaner tails after regrinding to 97% minus 200 mesh are sent to the copper scavenger cleaner circuit. This cleaner circuit consists of eight Denver #24 Sub A cells and three #36 agitair cells. The Denver cells are normally operated four, two and two to give a 24% copper concentrate product in three stages of cleaning. The agitair concentrates are recycled via the regrind mill while the scavenger cleaner tails is sent to the copper scavenger cells.

Lime slurry is added to copper flotation circuits to maintain a PH of 9.5 in the roughing and 10. in cleaning. Of the Z 200 collector added about 60% is to the copper conditioner and the remainder to the scavenger feed. The R 404 is split evenly between the primary and regrind ball mills while R 208 is split evenly between conditioner and scavenger feed. Depressants for sphalerite such as zinc sulphate have been tried without success.

Sphalerite is floated from the copper flotation tails after a 20 minute conditioning period. Zinc flotation is carried out in three banks of eight No.36 agitair cells. Normally these are arranged twelve cells for zinc roughing and twelve cells for zinc scavenging. The zinc roughers in turn are laundered in groups of four cells such that concentrates may be advanced in the cleaning stages as illustrated in the appended flowsheet. Zinc scavenger concentrates are returned to the zinc flotation feed. A rougher concentrate grade of 35% zinc is desirable for cleaner feed.

Zinc rougher concentrates are cleaned in four stages. The first stage has four No. 48 agitair cells, the second has six No.36 cells while the third and fourth stages have four No.36 agitair cells each. Concentrates proceed upstream to the final product from the fourth stage while tailings are directed downstream to first stage tailings which are returned to the zinc rougher feed.

Lime slurry is added to the No.1 conditioner to maintain a zinc roughing ph of 11.0 and to cleaner feed to maintain a ph of 11.5. Copper sulphate is added to the No.2 conditioner and potassium amyl xanthate to the No.3 conditioner, third group of roughers and final eight cell scavenger bank in the proportion of 40, 50 and 10 percent respectively.

CONCENTRATE DEWATERING

Copper and zinc concentrates are thickened in respective 20' diameter thickeners and each filtered on 5 disc by 6' Peterson filters fitted with filtex 90 bags. Vacuum is maintained at 24 inches by a Nash 2002 vacuum pump. Copper concentrates are filtered to 10% moisture content and trucked weekly to railhead for shipment to the Noranda smelter. Zinc concentrates at 12% moisture are further dried to 6.5%

moisture content in a 3' diameter by 20' Lockhead-Haggerty dryer. Dried zinc concentrates are stored under cover at the mine and shipped periodically in Fabrene covered Gondolas to Quebec city for overseas transport.

TAILINGS DISPOSAL

Mill tailings are pumped approximately 3000 feet through a 6" diameter thinwall black iron pipeline to the tailings impounding area. On site material is used for tailings dam walls, tailings are discharged around the perimeter of the area from 1" diameter spigots in the tailings line. Decanted water is stored in a secondary pond adjacent to the tailings area from which all the clarified water is returned to the mill for reuse. Make-up water is obtained from a local source.

MILL CONTROL

Lime slurry is pumped from make-up tanks through a loop and additions to the circuits are manually adjusted through pneumatic valves operated by percentage timers. Copper sulphate is batch prepared and pumped to a feed tank from which it is metered by pump to the zinc circuit. Other reagents are fed by Clarkson feeders and flowrators.

An Inax XRF analyzer, in operation since start-up of operations, has proven to be most valuable in control of the flotation operation. Analysis is available in minutes and a routine suite of samples are recorded on a two hour basis. The instrument is interfaced with a tektronic programmable calculator which also allows use for mill accounting.

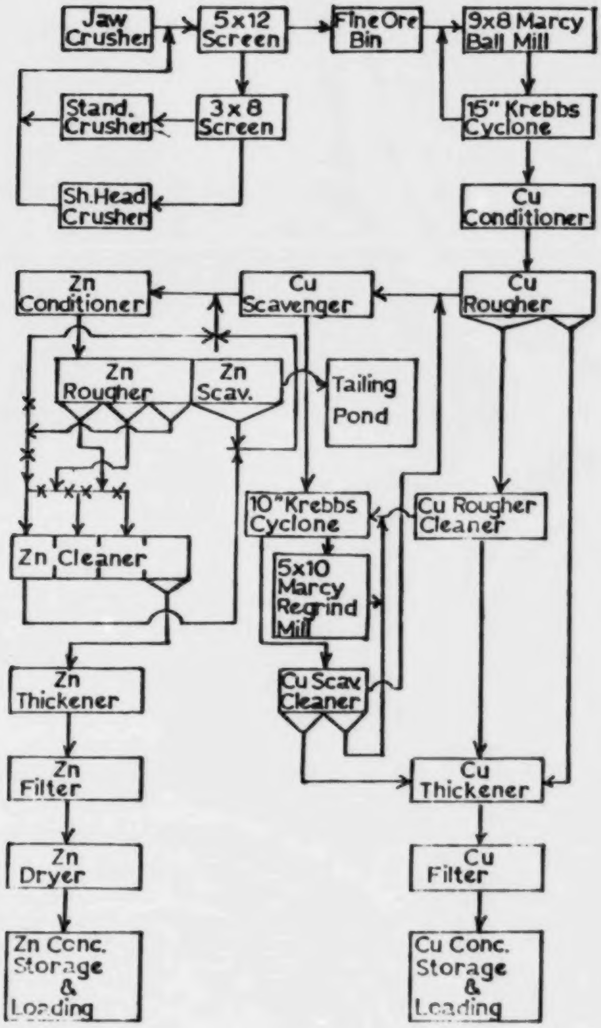

Jaw Crusher
5x12 Screen
Fine Ore Bin
9x8 Marcy Ball Mill
Stand. Crusher
3x8 Screen
15" Krebbs Cyclone
Sh.Head Crusher
Cu Conditioner
Zn Conditioner
Cu Scavenger
Cu Rougher
Zn Rougher
Zn Scav.
Tailing Pond
10" Krebbs Cyclone
Cu Rougher Cleaner
Zn Cleaner
5x10 Marcy Regrind Mill
Cu Scav. Cleaner
Zn Thickener
Cu Thickener
Zn Filter
Cu Filter
Zn Dryer
Zn Conc. Storage & Loading
Cu Conc. Storage & Loading

Figure 1 - Lemoine Mines Limited Mill Flowsheet.

Table 1 - Reagent Consumption
(Lbs/Ton Milled)
Jan - Aug 1978

Reagent	Circuit	
	Copper	Zinc
Lime	2.973	0,991
Z-200	0.073	
AF-208	0.208	
AF-404	0.084	
Copper Sulfate		1.957
R-355		0.054
D-250		0.016

Grinding Media	Lb/Ton
3" Forged Steel Balls	1.755
1" Forged Steel Balls	0.699

Table 2 - Metallurgical Performance Jan. - Aug. 1978

Product	Assays				Distribution			
	Au Ozs / Ton	Ag Ozs / Ton	Cu %	Zn %	Au %	Ag %	Cu %	Zn %
Mill Feed	0.161	2.82	5.03	11.02	100.	100.	100.	100.
Cu Conc.	0.655	11.66	25.04	7.19	77.89	78.86	95.05	12.45
Zn Conc.	0.121	1.71	0.59	54.06	12.32	9.92	1.91	80.13
Tailling	0.024	0.49	0.24	1.27	9.79	11.22	3.04	7,42

Table 3 - Man Power		
	Number	Schedule
Mill Superintendant	1	5 & 2
Lab. Technicians	2	5 & 2
Assayer	1	5 & 2
Shift Bosses	4	21 & 7
Flotation Operators	4	21 & 7
Ball Mill Operators	4	21 & 7
Helpers	4	21 & 7
Crusher Men	2	5 & 2
Mechanics	2	5 & 2
Total	24	

Table 4 - Operatin Cost	$/ton
Crushing	0.49
Grinding	0.94
Flotation & Filtration	2.07
Conc. Drying	0.32
Conc. Handling	0.16
Tailings Disposal	0.28
Benefits	0.70
Total	4.96

Discussion of Paper by G. Ladouceur

G. Johnston: Do you have any problems caused by the recycle water ?

Reply: For two years we had problems that could have been attributed to the water but in the following year we had acceptable results. We have about 1 1/2 months of holding time before it enters the mill. We had some problems with precipitation the water that blackened up the lines but we added a secondary settling pond and the problem disappeared.

A. Mai: Could you elaborate on the spigoting lines ?

Reply: It is of 6 inch mild steel and has been in service for three years. We grind to 80% minus 200 mesh. The spigoting is around the perimeter in summer. Most of the coarse is kept to the bank to allow better percolation. The lines are of ABS with holes drilled in it.

A. Williams: What are the openings on the rod deck ? How do the liners stand up to the 3 inch grinding balls in the ball mill ?

Reply: In a closed circuit mill there is a tendency for the liners to break. In our case we have a second hand mill that was cracked and we had to put in rubber liners to protect it. The cost for the first set of liners was 12 1/2 cents a ton. The second set cost 18 cents a ton. The opening on the rod deck is 3/8 by 6 inches. The opening on the second screen is 1 by 6 inches.

Paper No. 13

AN APPROACH TO ENERGY MANAGEMENT
AT
FALCONBRIDGE NICKEL MINES LIMITED

D. M. Doyle
Energy Conservation Co-ordinator
Falconbridge Nickel Mines Limited
Falconbridge, Ontario

ABSTRACT

A study of the various components contributing to energy costs and their relationships to generally trended operating data, such as tonnage, operating costs, etc. over progressively shorter time periods provides information for an energy audit. A study of annual data gives some insight into the importance of various energy components. As the time frame of the study progresses through monthly, daily and finally, into quarter hourly data, the variations become more evident and most important, strategies develop to minimize energy costs. The results of an audit conducted during the summer of 1978 at Wesfrob Mines are reported to illustrate this important phase, prerequisite to a meaningful energy cost avoidance program.

The balance of the paper deals with the energy conservation program at Falconbridge Nickel Mines Sudbury Operations, the metering and control equipment, the commitment, the new energy accounting practices and lastly, the policies. Examples of successful energy conservation endeavors include automatic peak electrical load control, waste heat recovery from mine air and compressor cooling water, energy reductions resulting from lower building temperatures and curtailment of idling loads. The use of pulse meters to measure the flow of gas, steam, compressed air and electrical energy, with their signals interfaced to process computers, allows monthly, daily and if necessary, immediate reporting of energy and energy per unit of production information. The program so reported at Falconbridge's Sudbury Operations is credited with an annual cost avoidance of .9 million dollars in a 14.0 million dollar annual energy account.

An Approach To Energy Management at Falconbridge Nickel Mines

The energy conservation efforts within most of the Falconbridge Nickel Mine group of companies, have to this date, not required major capital expenditures but rather have been directed towards the following:

1. Developing a knowledge of the significant energy consuming loads.
2. Installation of metering and controlling equipment.
3. Obtaining commitments to reduce unit energy consumption.
4. Revising accounting practices to encourage conservation.
5. Development of an ongoing co-ordinated policy.

The findings of the energy audit conducted at Wesfrob Mines Ltd., British Columbia, during the summer of 1978, are used as an example of a study made to develop a knowledge of the significant energy consuming loads. The audit provides the framework for the subsequent energy conservation programs. Examples of the energy conservation programs which are being implemented as a result of the audit conducted at Falconbridge's Sudbury Operations serve to illustrate the remaining four criteria necessary for an energy cost avoidance program.

Item 1 - An Energy Audit of Wesfrob Mines Ltd.

A study of annual production, energy and cost provides a cursory insight into the relative importance of energy costs to total costs; trends in tonnage, costs and fuel prices; the labour and material costs of energy; and the relative importance of various energy fuels. As exemplified by Figures 1, 2 and 3, a change from open pit to underground mining, resulted in a downward trend in the tons of ore and waste mined tons of ore milled onsite operating and energy costs. The cost of energy relative to onsite operating costs and to tonnage mined rose due to price increases in fuels and treatment of lower annual tonnage.

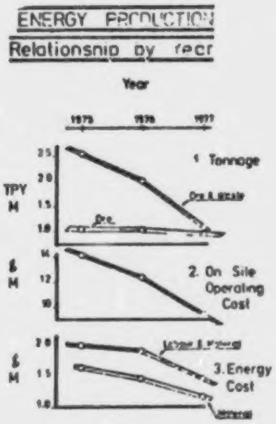


Figure 1

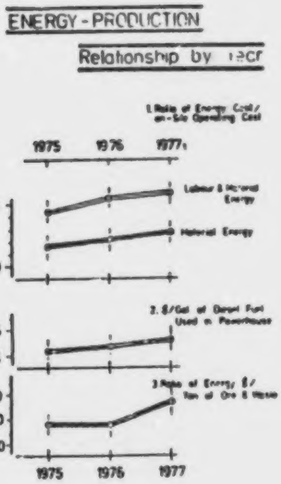


Figure 2

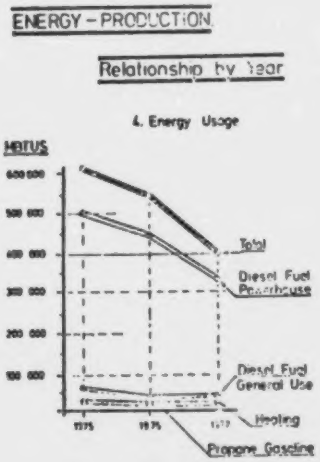


Figure 3

The energy consumed in the operation of the diesel generating plant, accounts for the major energy usage. The thrust of the energy audit was therefore directed towards an understanding of the electrical generation and electrical energy usage.

A study of monthly production energy and cost data introduces an awareness of wide month to month variations. The example on Figure 4 shows the varying ratios of energy consumption per unit of production. The efficiency of the diesel generating plant and the energy content of the fuel are combined to yield the conversion factor of 1 KWh = 10,000 BTU. Data is omitted for the periods in late 1977, as the operations were temporarily suspended.

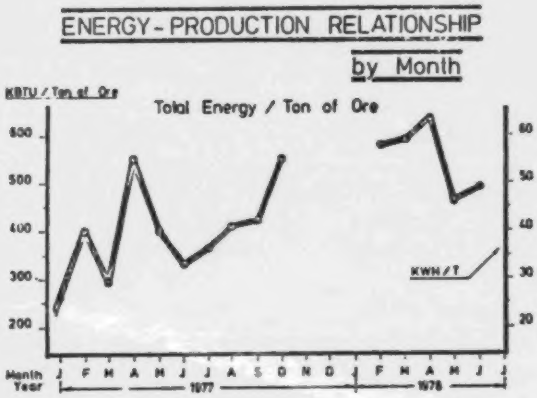


Figure 4

As the time frame of the energy study is reduced to a daily period, and in this example to a measure of electrical demand every 4 hours, the wide variation in energy usage became more apparent but more important, explanation for and solutions to the problems are identified. The problem is excessive demand, requiring peak control strategies and excessive energy consumption, requiring curtailment of idling loads. The steady decrease in the weekend demand over the month resulted from conservation efforts.

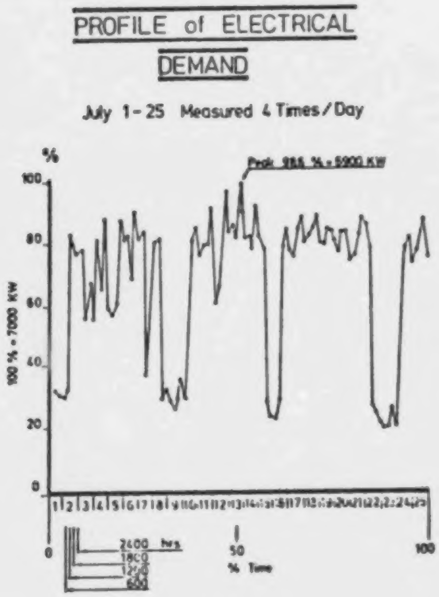
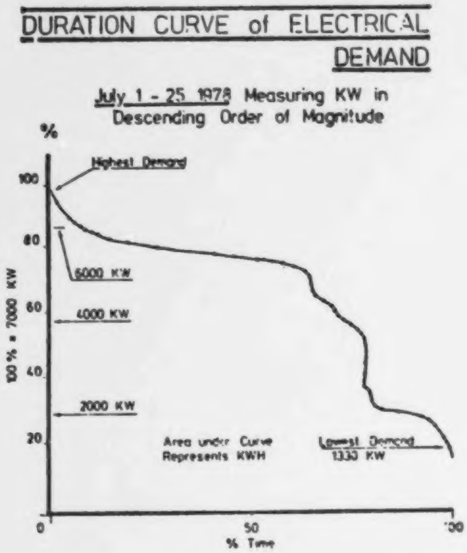


Figure 5

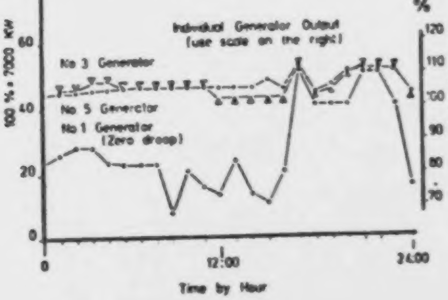
The electrical demand data from the previous figure, shown in descending order of magnitude, (Figure 6), suggests that the operating load during production periods and weekends should be 6000 KW and 1400 KW respectively.


DURATION CURVE of ELECTRICAL DEMAND

July 1 - 25 1978 Measuring KW in Descending Order of Magnitude

Figure 6

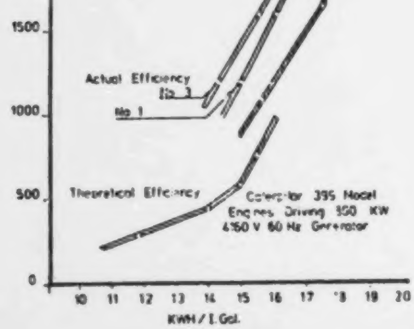
A study of the electrical demand over each hour of the peak demand day, on the system and from each generator, shown by Figure 7 suggests that the peak could be reduced and the sytem run at a steadier demand if some work was rescheduled from the 4-12 shift.


PROFILE of ELECTRICAL DEMAND

Peak Day July 13/78

Figure 7

Comparisons of the electrical energy generated and the fuel consumed by the Mirrlees diesel engines, provides a comparison between theoretical and actual operating efficiencies of the diesel generators. Optimum efficiency occurs when the generators operate at rated capacity.


EFFICIENCY CURVES

Comparing Generator Output (KW) to Fuel Efficiency (KWH / I.Gal.)

Figure 8

Comparisons between the amount of electrical energy consumed throughout the operation each day and the tonnage milled (Figure 9) shows variation in energy/ton, well beyond that expected from changes in ore hardness. It became apparent that base load energy (required to sustain basic electrical needs) and idling load energy (required to turn equipment), are sizeable contributors, which when added to production energy (size reduction, separation, etc.) equal total electrical energy requirements.

ENERGY - TONNAGE RELATIONSHIP

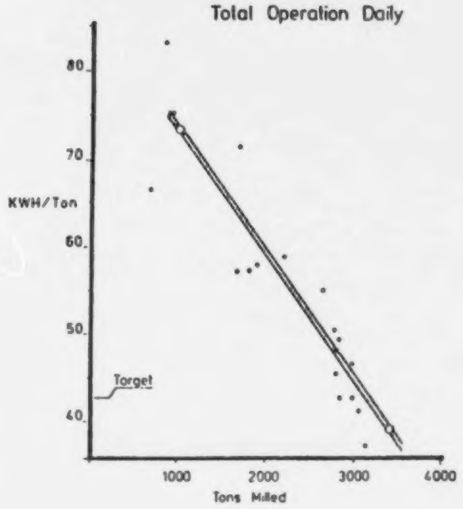


Figure 9

Comparison between the electrical energy consumed daily in the grinding and crushing processes and throughput rates (Figure 10) shows the relationships existing between TPH and KWh/T. Tonnage rate is to a large degree influenced by the inventory of ore. Attainable energy per ton targets can be met by either increasing the availability of ore or reducing the number of operating hours.

ENERGY - TONNAGE RATE

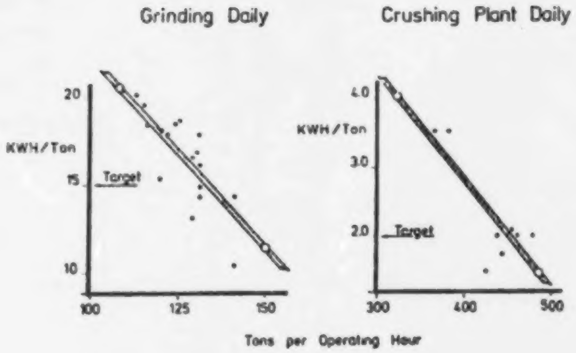


Figure 10

Measurements of the energy used at each cost centre, and at each of the major loads while idling and while operating under load allows one to i) identify and quantify the energy users by cost centre, ii) approximate the amount of energy used to sustain the operation, to turn the equipment and to produce useful work, and iii) to compare the relative importance of the energy consuming processes. For example: (Figure 11) shows the mill department consuming 76% of the total onsite energy. (Figure 12) shows that 60% of the energy used in the operation is to idle the crushers, conveyors, compressors, etc. while only 27% is used to perform useful work (i.e. size reduction particle separation) and that 13% of the energy used on a normal operating day is to sustain the basic electrical needs in the townsite, mine fans and mill pumps. (Figure 13) shows that grinding, flotation and air compressing account for the major energy consuming processes while lighting, an energy user demanding far more attention than necessary, accounts for less than 3% of the total normal operating day electrical load.

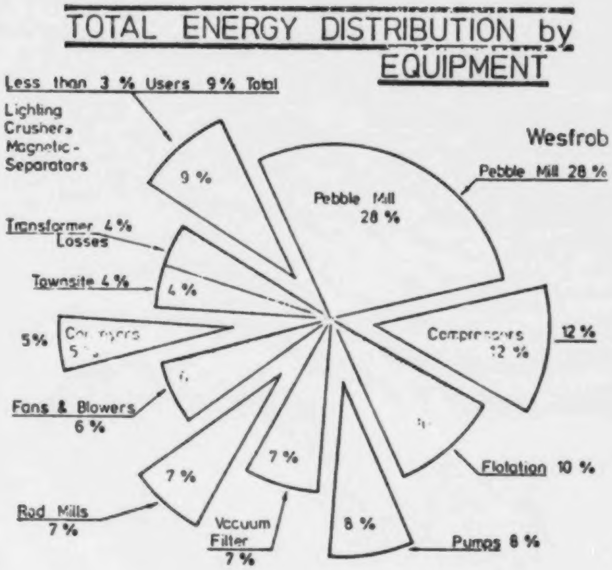


Figure 11

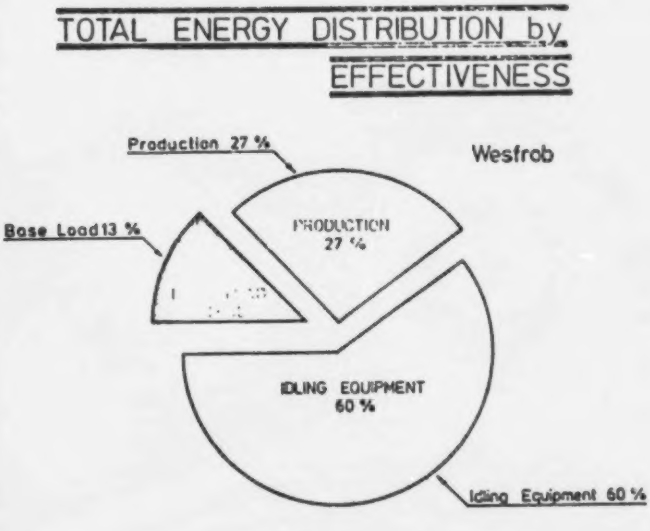


Figure 12

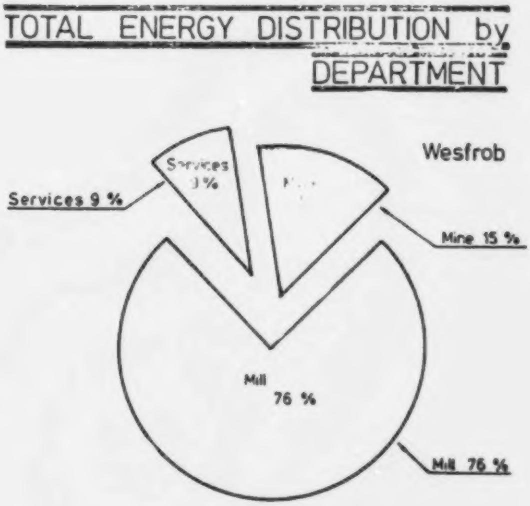


Figure 13

Items 2-5 - The Conservation Program At Falconbridge's Sudbury Operations

The findings of the energy audit conducted on the Sudbury Operations were naturally different from those at Wesfrob, however a similar approach was taken in order to gain the knowledge prerequisite to formulation of an effective energy conservation program. (Reference 1). The ongoing program being followed at the Sudbury Operations is highlighted in the following section.

A Heating Energy

Heating costs represent 18% of total energy costs and one example (Figure 14) of a conservation effort in this regard includes the use of shell and tube type heat exchangers to recover heat from compressor cooling water and finned heat exchanger to recover waste mine air heat, transferring some 26 M BTUh at Strathcona to a glycol solution and in turn, heating mine intake air.

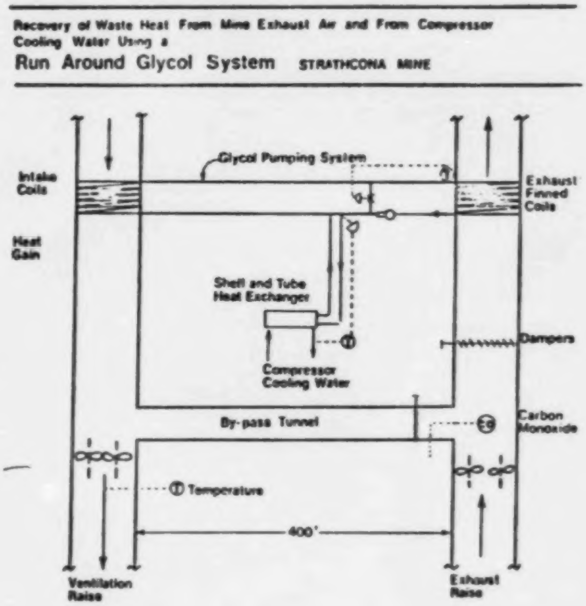

Recovery of Waste Heat From Mine Exhaust Air and From Compressor Cooling Water Using a
Run Around Glycol System STRATHCONA MINE
Glycol Pumping System
Intake Coils
Heat Gain
Exhaust Finned Coils
Shell and Tube Heat Exchanger
Dampers
Compressor Cooling Water
By-pass Tunnel
Carbon Monoxide
Temperature
400'
Ventilation Raise
Exhaust Raise

Figure 14

B Electrical Energy

In the 1980's, electrical energy will account for 78% of our energy costs. This fact has therefore influenced our program of cost avoidance through peak electrical control and electrical energy reduction.

Since 1972, we have used load sensitive current relays to restrict some 5000 KW of electrical demand from water heaters, mine pumps, hoists, and crushers, whenever the total electrical demand climbed to predetermined levels.
Figure 15 shows the variation in electrical demand in our Onaping Area over a number of hours, before and after installation of the peak sharing control system.

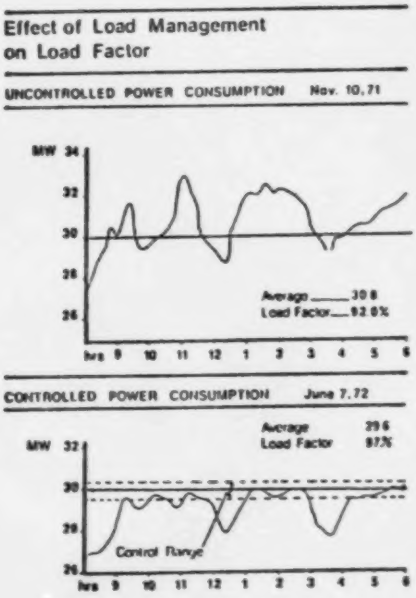


Figure 15

Underground storage of compressed air for peak electrical control is being considered in a number of mines. Our 3/4 mile long air receiver at the Falconbridge Mine, provided us with the confidence and experience for further development of this concept. Figure 16 shows a record of cfm of compressed air entering the receiver for storage, and exiting the receiver to supplement that produced by the compressors during brief periods of high air requirements. The operation of compressors during off peak electrical periods for storage of compressed air may some day make a sizeable reduction in mines electrical costs.

Air Receiver
24 Hour Record of Air Flow

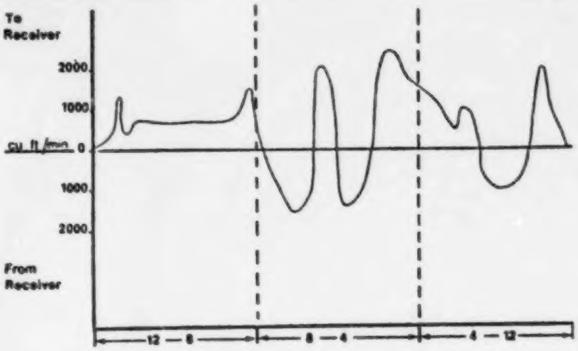


Figure 16

Studies of the electrical load for each 15 minute billing interval at each of our properties has shown patterns exemplified by Figure 17, showing excessive peaks, the target peak of 98% load factor measured on a shift basis, waste energy, and minimum base loads required to sustain the operation.

Daily Electrical Load Profile
Lockerby

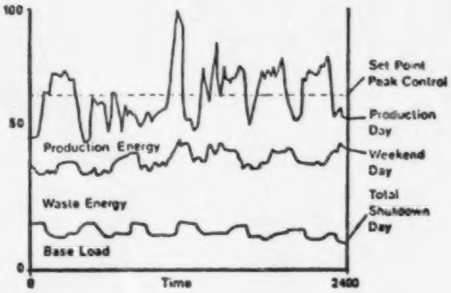


Figure 17

Studies of Duration Profiles produced from the demand data at each billing area (Figure 18) have shown the potential savings to be realized by peak control and conservation and this has led to the installation of peak load control devices at each billing area.

Monthly Load Duration Profile Summary
Lockerby 76 and Ideal Profile

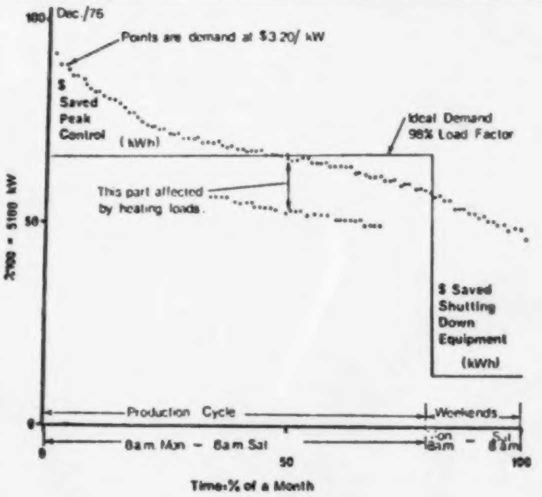


Figure 18

The installation of a modest peak load control system and a commitment from operators to reduce energy consumption at one of our inactive properties, is an example where $9,000 investment resulted in a $110,000 annual cost avoidance.

Figure 19 shows the electrical demand profile every quarter hour of a peak day, before and after the conservation program. Figure 20 shows the duration profile of electrical demand for 2 week periods, before and after control. The comparables were made at similar times of year and level of activities.

DAILY LOAD PROFILES OF
ELECTRICAL DEMAND
AT LOCKERBY MINE

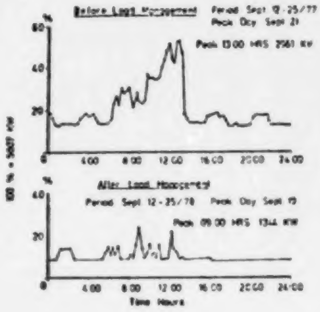


Figure 19

DURATION PROFILE OF
ELECTRICAL DEMAND
AT LOCKERBY MINE

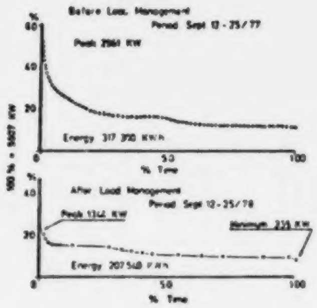


Figure 20

A duration profile of the electrical demand over one month on a 2600 hp centrifugal compressor (Figure 21), illustrates the energy savings resulting from stopping this motor on weekends. We now routinely shutdown compressors, pumps, fans, heaters, which formerly operated continuously irrespective of operating requirements.

Order of Magnitude Curve Centrifugal Compressor

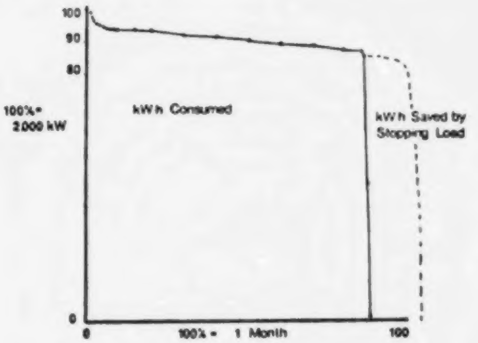


Figure 21

A computer control concept applicable to peak load control of heaters, reciprocating compressors, DC hoists and in this example, mine water pumps, (Figure 22) deserves consideration. Two additional relays are inserted into the sump level pump control and are activated by the computer whenever the area electrical demand is in either a "peak" or "valley" condition. Water is stored when electrical demand is in a peak, and water is pumped when electrical demand is in a valley.

COMPUTER CONTROL of Mine Water Supply

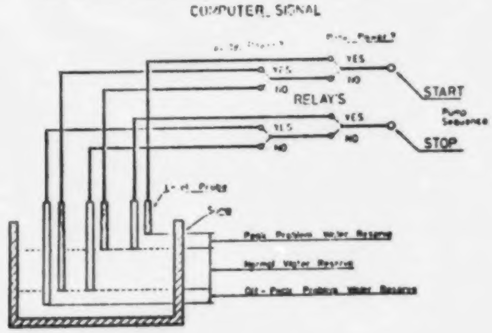
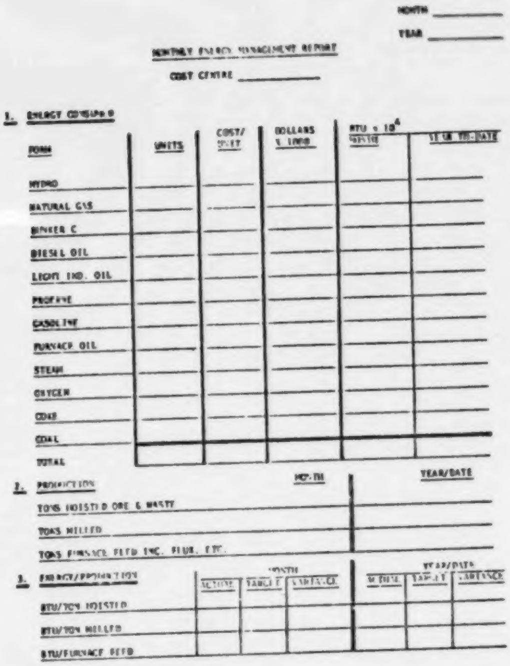


Figure 22

C Energy Accounting

Regular and formal accounting of energy consumption to each cost centre, in the Sudbury Operations is to become standard practice. We have been installing meters to measure the amount of electrical energy, compressed air, steam, oil, and gas to each cost centre and each month report the amount of energy, the cost, the energy/ton ratios and variances from target figures. The onus to conserve and to reduce consumption below comparable operations becomes the effectiveness area of the individual plant superintendent.



MONTH __________

YEAR __________

MONTHLY ENERGY MANAGEMENT REPORT

COST CENTRE __________

1. ENERGY CONSUMED

FORM	UNITS	COST/UNIT	DOLLARS X 1000	BTU x 10^6 MONTH	YEAR TO-DATE
HYDRO					
NATURAL GAS					
BUNKER C					
DIESEL OIL					
LIGHT IND. OIL					
PROPANE					
GASOLINE					
FURNACE OIL					
STEAM					
OXYGEN					
COKE					
COAL					
TOTAL					

2. PRODUCTION

	MONTH	YEAR/DATE
TONS HOISTED ORE & WASTE		
TONS MILLED		
TONS FURNACE FEED INC. FLUX, ETC.		

3. ENERGY/PRODUCTION

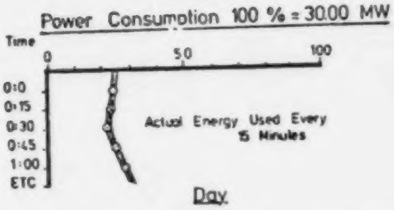
	MONTH			YEAR/DATE		
	ACTUAL	TARGET	VARIANCE	ACTUAL	TARGET	VARIANCE
BTU/TON HOISTED						
BTU/TON MILLED						
BTU/FURNACE FEED						

Figure 23

The pulse type meter was selected for the measurement of compressed air, steam and electrical energy, and continuous measurement of gas and oil will be added in the future. The amount of energy used in every 15 minute discreet time frame is recorded in our process computers. Figure 24 illustrates a daily electrical energy report. The bottom line on all reports show the dollar cost of energy.

DAILY LOAD PROFILE

Sept 24/78



Day

	12-8	8-4	4-12	TOTAL
Peak Time	4:45	15:45	20:15	20:15
Peak KW	8400	8640	8880	8880
Peak 09 KVA	8300	8400	8500	8500
KWH	60360	60720	63300	184380
Load Factor	89.8	87.9	89.1	86.5

MONTH / DAY

MWH	2981.0	2995.4	3053.4	9029.8
KW Peak				24.000
Load Factor	64.7	66.5	66.8	66.0
Energy $				92104
Demand $				107040
TOTAL $				199144

Peak Load Occured 17:45 on Sept 7 at 24 MW

Figure 24

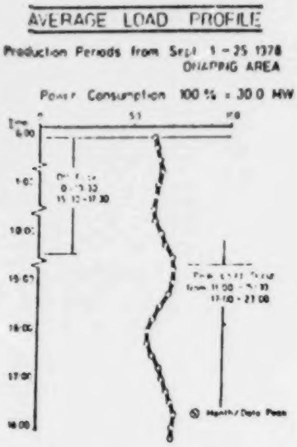


Figure 25

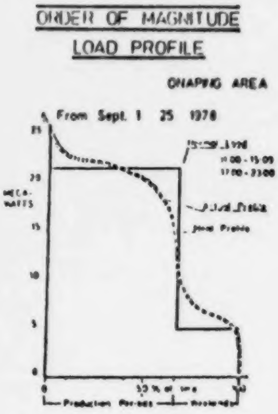
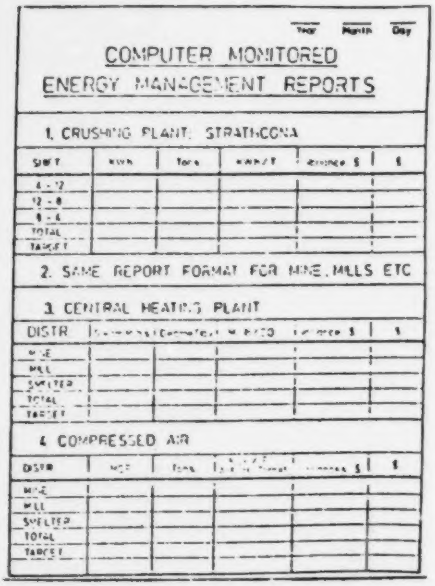


Figure 26

On the 26th and 1st day of each month, the computers report the average demand during production period, the peak demand (Figure 25), and the duration profile (Figure 26). This information is used to determine the target set points for the following month, to illustrate the hill and valley hours which may be smoothed by rescheduling, and most important, to obtain commitments for demand and energy reduction. Computer monitoring of energy consumption enables us to alarm when consumption exceeds attainable target levels and report, as exemplified by Figure 27, the ratios of energy to tonnage and lost dollars resulting from operating the plant at less than optimum rates.



Year Month Day

COMPUTER MONITORED
ENERGY MANAGEMENT REPORTS

1. CRUSHING PLANT: STRATHCONA

SHIFT	KWH	Tons	KWH/T	[illegible] $	$
4 - 12					
12 - 8					
8 - 4					
TOTAL					
TARGET					

2. SAME REPORT FORMAT FOR MINE, MILLS ETC

3. CENTRAL HEATING PLANT

DISTR	[illegible]	[illegible]	[illegible] RATIO	[illegible] $	$
MINE					
MILL					
SMELTER					
TOTAL					
TARGET					

4. COMPRESSED AIR

DISTR	MCF	Tons	[illegible]	[illegible] $	$
MINE					
MILL					
SMELTER					
TOTAL					
TARGET					

Figure 27

Peak Electrical Load Control

The electrical peak load controllers and computer software programs operational on all our billing centres (Figure 28) measure the energy accumulated within each 15 minute billing interval, measure the rate of change, predict the demand at the end of each billing interval and add or shed loads to achieve a target demand. This phase of the program will continue, especially with "time of day" billing considerations, however we are very pleased with the results to-date.

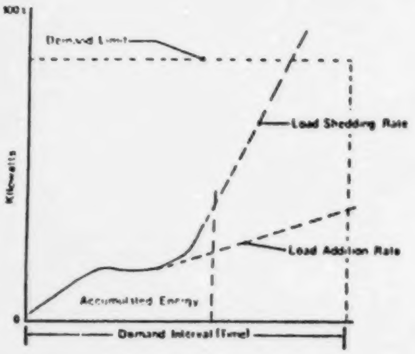


Figure 28

Conclusions

Accomplishments in the Sudbury Operations to-date (Figure 29) represent an annual cost avoidance of .9 million dollars, in our total 14-15 million dollar energy accounts.

ENERGY
CONSERVATION ACCOMPLISHMENTS

1 AWARENESS
2 PEAK ELECTRICAL CONTROL
3 GLYCOL RUN-AROUND HEAT RECOVERY
4 METERING
5 HEATING ENERGY REDUCTION
6 ELECTRICAL ENERGY REDUCTION

Capital requirements have been modest and when required payback has been less than 2 years.

The audit provided an awareness of the problem, it qualified and quantified our energy consuming processes and led to a very active program involving management, energy task force officers, and through a video presentation, all of our employees.

Heating costs are being reduced by operation of our buildings with less air infiltration and at lower temperatures. The glycol run around systems at our Lockerby and Strathcona Mines recover heat from compressors and mine exhaust air for preheating mine air.

The peak load controller installed in 1972 was the first in the Canadian Mining Industry and has guided us into better control using pulse metering. Whenever the production cycle allows, large electrical loads are shutdown.

The pulse type metering of all forms of energy to each cost centre will provide continuous auditing of energy and continuous comparison of energy consumption per unit of production, to known attainable results.

I want to acknowledge the contributions of numerous employees of Falconbridge Nickel Mines.

References:

Doyle, Donald M. "Energy Conservation of Falconbridge Nickel Mines Limited", Canadian Mineral Processors Proceedings - 10th Annual Meeting 1978.

Discussion of Paper by D. Doyle

E. Joe:	Are there any means for the equipment to shut itself off if the load ceases to exist ? Are you going to include the detailed procedure in the Proceedings to obtain this type of data.
Reply:	In the case of the large electrical load shutting down when the feed stops, we do not have it automatic but we are close to it. The details will be in the Proceedings. However, every mill superintendent will have to do his own. Sooner or later someone in you own company or in a government will be asking for your BTU per ton and your committments.

Paper No. 14

OPERATION AND MAINTENANCE OF MINERAL PROCESSING PLANTS

by: A.L. Mular

On behalf of Bob Bruce and myself I would like to begin by stressing some key points.

First, most of us know that the Canadian Mineral Processors was formed in 1969 and is an offshoot of the Canadian Gold Metallurgists. I attended that first meeting in 1969, and I haven't missed a meeting since! Pretty obvious that CMP serves the best booze! At any rate, we know that our meetings are supported by CANMET. Camsell Hall is always at our disposal; Bob Bruce acts as permanent secretary pretty much free of charge; we receive a financial grant from CANMET. This brings me to a second point.

Second. the Canadian Mineral Processors do not enjoy a free ride in this relationship. CMP provides a genuine service to the mining industry, to government, and to universities. For the industry, CMP serves as a forum for the rapid dissemination of new ideas and new technology developed at remote locations throughout Canada. New developments outside Canada are introduced by invited speakers at our meetings. The effect is to help reduce operating costs and increase efficiency.

For government, problem areas and research priorities can be isolated through mutual discussions. Direct communication at our meetings can help CANMET to temper its goals and modify its national priorities.

For universities, CMP provides a wealth of operating experience and practical know how. The yearly proceedings are a major current reference source for those of us who teach mineral processing. Admittedly, the proceedings are published from CMP coffers. However, I note that our fees are up this year, which brings me to a third point.

Third, we face some potential problems of a financial and political nature. From a financial view, we may lose our CANMET grant and other financial good will simply because of spending constraints. Our fees are expected to climb each year thereafter.

From a political view, we know that the metallurgical society of CIM is anxious to incorporate CMP as a section. The pros and cons of such a merger have been described in a recent mailing. In the final analysis, the political and financial problems are interrelated. There are several alternative solutions to both problems, which brings me to a fourth point.

Fourth, some solutions to the political problem are: (1) join the METSOC, (2) affiliate in a loose way, (3) just ignore them. Frankly, I do not know what the decision, based on the recent poll, will be.

Of the possible solutions to a potential financial problem, we suggest that CMP undertake the task of writing a textbook.

Now I've talked with several of you about a textbook, so I've done some homework. Several questions arise, namely, what, why, who and how!

The answer to what is:

Operation and Maintenance of Mineral Processing Plants

This is the one subject that you know best--you are the experts.

The answer to why is twofold:

a) We estimate that a text of this sort will be a best seller. My guess is that within 3 years, a minimum of 5000 copies can be sold. At the fair price of $30 per copy you should net of the order of $70,000 and I think my estimate is conservative. This amount goes a long way towards solving the financial problem. In fact, royalties will accrue for many years after.

b) Again the CMP will perform a badly-needed service.
   1. For industry, we could have a training text for operators, for engineers not trained in mineral processing, and for lawyers who seem to pack our our boards of directors.
   2. For government, we expose our daily know how and experience to the applied research scientist who then might direct his efforts along more practical lines of endeavor.
   3. For universities, well we graduate young men who feel somewhat lost during their first few months of exposure to a production environment. A text on operation and maintenance will certainly be a marvelous orientation.

Now I wish to stress that this text will not replace the principles of unit operations and unit processes that we teach at 3rd and 4th year levels. Rather, such a text will help us develop courses in plant design and practice.

Now, who will do it? Well, we need co-editors, a general editorial committee, someone to help with drafting, layout and indexing, and a publisher.

We contacted two well known and respected Canadian mineral processors to determine whether they might act as co-editors; one from the east and one from the west. The one from the east was too busy to undertake such a task; the one from the west thought it was a great idea but he had several reasons why he didn't want to take the job. This led Bob and I to conclude that the co-editors had to have plenty of spare time! Academics and government employees have plenty of spare time, so Bob and I volunteered for the job.

For a general editorial committee, we have several who have stated that they would be interested in the project.

We have persons who might be willing to help with drafting, layout and generating an index, and we have some ideas on publishing the text.

Here is how we might get started:

CMP approves idea, content, staff and schedule.

Co-Editors: Mular; university view, no ax to grind, more spare time.
Bruce; government view, no ax to grind, more spare time.
Ensure text is grammatically correct and settle differences.

Editorial: Backbone of text; assigned chapter responsibilities; delegate accordingly.

Drafting,: Pickett for this purpose. Might pay him a small sum for his services.

Publisher: Printing firm who will do the job at a flat rate using the printing procedures employed by AIME. (Cite plant design text)

Text description

So, in brief summary, we think that the Canadian Mineral Processors should help themselves by putting a textbook together. Bob and I are prepared to donate our time at no cost and I hope you are. We think you will have a best seller on your hands for years to come.

Suggested Content and Layout

A - Title page

B - Cover page

C - Forward

D - Preface

Chapter	Title
1	Organization of Typical Mine-Mill Complexes
2	Inventory, Accounting and Forecasting Procedures
3	Typical Tables of Organization For Processing Plants
4	Mine-Mill Safety Programs
5	Basic Structural and Orgainzation of Labour Unions
6	Organization of Maintenance Departments
7	Primary Crusher/Screening Plants
8	Secondary Crushing/Screening Plants
9	Dust Collection Systems
10	Primary Grinding Circuits
11	Semi-Autogeneous Grinding Circuits
12	Re-Grind Circuits
13	Coal Screening Circuits
14	Tabling and Jigging Circuits
15	Heavy Media Separation Circuits
16	Humphrey Spiral Circuits
17	Electrostatic Separation Circuits

Chapter	Title
18	Magnetic Separation Circuits
19	Thickener Circuits
20	Filtration Circuits
21	Rougher/Scavenger Flotation Circuits
22	Cleaner/ Recleaner Circuits
23	Tar Sands Separation Circuits
24	Cyanidation Circuits
25	Ion Exchange/Solvent Extraction Circuits
26	Tailings Disposal Systems
27	Concentrate Drying, Storage and Transportation
28	Cu-Mo Concentrator
29	Pb-Zn Concentrator
30	Cu-Ni Concentrator
31	Au Concentrator
32	Potash Concentrator
33	Asbestos Concentrator
34	Tungsten Concentrator
35	Uranium Concentrator
36	Iron Ore Concentrator
37	Coal Washery
38	Tar Sands Concentrator
Index	

Chapter 7 to 27 Inclusive Contain:

A. Typical Flow Diagrams

B. Process units involved (e.g. secondary, tertiary crushers; screens; bins; conveyors; feeders; etc.) and corresponding purpose.

C. Operation
   1. Variables controlled and why.
   2. Variables manipulated to control and why.
   3. Disturbance variables not regulated and how to compensate, Hazards (i.e. tramp steel) and how to deal with them.
   4. Description of operator responsibilities
   5. Operating times (length of shifts, etc.)
   6. The typical shift (operator inspections, etc.)

D. Maintenance
   1. What to service on a daily basis; down times; type of inspections
   2. What to service on long term basis; down times; types of inspections
   3. Preventive maintenance
   4. Support facilities required
   5. Job descriptions for maintenace personnel required for tasks
   6. The typical maintenance routine (daily and long term)

E. Instrumentation in Common usage

F. Nature of traning Programs

Chapters 28 to 38 Inclusive contain :

A. Plant flow chart; major process units

B. Organization

C. Routine features of daily operation

D. Special features of operation

E. Types of supplies; consumption tables; etc.

F. Interactions with community

Discussion of Paper by A. Mular

I. Itzkovitch: I think the request is premature. A business plan is required showing the costs and revenue expected.

Reply: I don't agree with you. All I am interested in at this time is agreement in principle.

D. Livingstone: Is there any other publication that parallels this ?

Reply: No. Not in the world.

E. Joe: Mineral Processing Plants has recently been published and it took 10 years in the making.

Reply: This is not along the same lines.

K. Konigsmann: Basically your idea is good but the time allowed for thinking about it is too short in order to make such a decision. I suggest that we discuss it during the evening.

Reply: Agreed.

Paper No. 15

INDUSTRIAL APPLICATION OF MATHEMATICAL MODELLING TO THE TREATMENT OF FINE-GRAINED SULPHIDE ORES

by

I.B. Klymowsky* and W.H. Cameron**

SUMMARY

The paper presents an overview of the work being done at CANMET on the modelling of grinding and classification. Models for a ball mill and hydrocyclone are described, and new analytical expressions for the model parameters are introduced. Results of laboratory, pilot plant and industrial experiments on grinding and classification are reviewed. The efficiency in grinding is shown to be constant for these ores under a wide range of operating conditions, and classification is identified as the area with the greatest influence on the final grind.

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INTRODUCTION

The particular ores being studied are the fine-grained Zn-Pb-Cu ores of the New Brunswick area. These ores represent a major, yet largely unexploited, resource of Zn, Pb, Cu, and Ag. Their treatment demands complex beneficiation procedures. Recoveries of metal values from these ores presently range from 50% to 80%. The overall objective is to raise these metal recoveries above 90%.

The approach has been to concentrate first on the grinding and classification of these ores:

1) to try to implement a much finer grind than that presently being managed in practice and,

2) to reduce the formation of slimes

The present grind, of 65-70% minus 37 μm (400 mesh), does not provide sufficient liberation to achieve 90% recovery. A much finer grind, approximately 90-100% minus 37 μm (400 mesh) is indicated.

Slimes account for a substantial proportion of the losses in flotation. A finer grind will require more attention to factors contributing to the formation of slimes, particularly slimes of the valuable minerals. One of these factors is poor classification, which promotes the formation of slimes by recirculating fine liberated materials through the grinding mills. Another appears to be rod-milling, which contributes significantly more slimes to the final product than ball-milling. Fig. 1 gives the size distribution of products from an industrial grinding circuit to illustrate this point. The percentage of minus 2 μm material in the rod mill discharge approaches that of the final product; the percentage of minus 2 μm material in the ball mill discharge is very small.

Efforts have been made in the past to establish a finer grind in practice, however, none of these have been very successful. There appears to be some underlying reason for the difficulty in implementing a finer grind in practice. In order to find this reason and to help define a practical method of achieving a much finer grind, modelling has been investigated.

The processes of grinding and classification are very complex. Modelling provides a means of systematically analyzing the effects of the large number of variables involved, allows testwork to be done within the confines of a mineral processing laboratory, and offers a basis for scale-up to industrial practice.

The following describes the models adopted for a ball mill and hydrocyclone.

Models:

Figure 2 shows a typical grinding circuit, and some of the parameters associated with the models. The models selected for each of the unit operations provide a description of the process in terms of particle size; therefore, knowing the size distribution of the feed and the values of the model parameters for different operating conditions, the size distribution of the product can be calculated.

Grinding Model:

Grinding is a kinetic process which can be described by three model parameters: Breakage (B), Selection (S), and the Residence Time Distribution (RTD).

The Breakage function describes the size distribution which results when a particle of a certain size is broken. Normally a particle does not split in two, but produces a whole size distribution of smaller-sized particles. Figure 3 illustrates the size distribution which results from the breakage of material in the 1770-1650 μm (10-14 mesh) size range. The breakage function appears to take on a specific shape, which can be described by two Gaudin-Schuhmann distributions(1). The expression has been simplified in this report with a two-parameter equation.

$$Y= \quad B(x/x_{max})^{\alpha} + (1-B)\ (x/x_{max})^{10\alpha} \qquad \text{equation (1)}$$

where: α and B are parameters
x = particle size
x_{max} = top size

The Breakage function is generally held to be a characteristic of the ore, and independent of milling conditions. Furthermore, it can usually be normalized; that is, the same percentage passes through consecutive screens regardless of the top size. Some degree of judgement should be exercised with respect to this assumption: very fine material may break differently from coarse material.

The Selection function describes the rate at which particles in the different size fractions are selected for breakage. This function is a characteristic of the mill, the operating conditions, and the hardness of the ore. Figure 4 illustrates a selection function observed in grinding experiments in a 40 cm (16")-diameter mill. In this particular case, the selection function exhibits a maximum, at approximately 707 μm (20-28 mesh), indicating that the balls may not be large enough, or that the mill diameter may not be great enough to provide the balls with sufficient energy, to break some of the larger-sized particles. In other cases, the Selection function has been found to be a simple linear-shaped function when plotted on log-log paper[2,3]. A three-parameter equation has been adopted to describe the shape of the Selection function:

$$\log(S \,.\, 100) = a(\log x)^{b}\, e^{c(\log x)} \qquad \text{equation(2)}$$

where: a,b, and c are parameters
x= particle size

This expression has several useful properties: it can display a maximum, it can exhibit linearity, and it tends to level out in the very fine size-range where the probability for breakage is the same for all particles.

The purpose of these analytical expressions is to permit simultaneous determination of the Breakage and Selection functions using as few unknowns as possible.

The Selection function is the key function in the grinding model. Its behaviour under different operating conditions describes the effects of these conditions on grinding. The product of the Selection function and hold-up weight of material in a mill (usually referred to as the absolute rate of breakage) is a measure of grinding efficiency. It lends itself readily to the development of scale-up parameters for grinding mills. The product of the Selection function and residence time is a measure of the amount of breakage occuring in a mill.

In a batch mill, all material has the same residence time. However, in a continuous mill, some portions of the material short-circuit the mill and are discharged almost immediately, while other portions stay in the mill for longer periods of time. The Residence Time Distribution describes the length of time material stays in the mill. Figure 5 illustrates a Residence Time Distribution curve measured in a continuous mill using tracers.(5) The results are plotted in such a manner that the total area under the curve is 1. Therefore, the fraction of material with residence time longer than t is equal to the area under the curve between t and ∞.

The Residence Time Distribution curve can be described mathematically using the mean residence time and variance about the mean, or using a mixers-in-series model.(4) A 2.5 mixers-in-series with a small plug flow component (equal to one-tenth of the mean residence time) has been adopted to describe grinding in a continuous mill. This model agrees well with the experimental observations in Figure 5, and simplifies to some extent the mathematics used in the grinding model.

The concepts of Breakage, Selection and Residence Time Distribution provide a description of grinding in terms which are physically meaningful. The values of these parameters can be determined experimentally and related to practice. However, the mathematics involved in calculating the values of the model parameters are quite complex and therefore not presented in this paper.

Hydrocyclone Model:

The hydrocyclone is also described by three model parameters: the reduced-efficiency curve, the d_{50} size and the water-split coefficient.

The reduced-efficiency curve is a characteristic curve of the process which describes the theoretical optimum sharpness of separation that can be achieved.

The d_{50} size is the size at which 50% of the particles report to the overflow and 50% report to the underflow.

The water-split coefficient is the ratio of water in the underflow to water in the feed.

The basic concepts of the hydrocyclone are two fold:

1) A certain proportion of the feed to the hydrocyclone by-passes the classification process and reports directly to the underflow as unclassified material. This "by passing " accounts for the slimes appearing in the underflow.

2) The remaining material is classified, but the efficiency with which this material is classified is fixed by the nature of the material (its shape and specific gravity), and is entirely independent of operating conditions.

There are two ways of calculating the amount of material by-passed. One is to assume that the amount of material by-passing the classification process is in the same proportion to the feed as the water appearing in the underflow. In other words, the proportion by-passed is equal to the water-split coefficient. The other, is to assume that the proportion by-passed is equal to the ratio of the fines appearing in the underflow to the fines in the feed.

Figure 6 illustrates an actual efficiency curve calculated from the results of a sampling campaign conducted around a 50-cm cyclone. Each point represents the percentage of material in a particular size-fraction of the feed that appeared in the underflow. Approximately 50% of the fines appeared in the underflow leading to the conclusion that 50% of the feed by-passed the classification process entirely.

An efficiency curve for the material which was considered classified can be drawn by subtracting material that by-passed the process from the total amount of material in each size fraction of the feed and underflow. This "corrected" classification efficiency curve Y(cor.) is also shown in Fig. 6:

$$Y(\text{cor.}) = \frac{Y\text{actual} - Rf}{1 - Rf} \qquad \text{equation(3)}$$

The reduced efficiency curve is obtained by normalizing the corrected efficiency curve with respect to the corrected d_{50} size.

Application

The water-split coefficient and the corrected d_{50} size will vary with operating conditions, but the reduced-efficiency curve will remain constant for a given material. This fact is demonstrated in Fig. 7,8, and 9 which show reduced-efficiency curves obtained in 10 cm (4 in), 37.5 cm (15 in) and 50 cm (20 in) hydrocyclones under different operating conditions.

An analytical expression for the reduced-efficiency curve has been developed at CANMET:

$$Yr = \tfrac{1}{2}\left((1 - e^{-ax^{b}}) + e^{-ax^{-\beta}}\right) \qquad \text{equation (4)}$$

where $x = d/d_{50}$ (corrected)
and a, b and β are
parameters

This expression offers the flexibility of having different curvatures for the upper and lower portions of the curve. It is also useful for determining the corrected d_{50} size.

The model of the hydrocyclone is sufficiently simple to permit calculation of the model parameters using a hand-calculator. However, these calculations can be lengthy and require complete circuit balances, which are not often readily obtained. To promote wider use of the model, a simple materials balance technique for complete grinding circuits has been developed.(6) Referring to Fig. 2, the technique consists of:

1) determining the circulating load(CL) from screen analysis, using the formula:

$$CL = \frac{\Sigma\,(RMD - COF)\,(CUF - BMD)}{\Sigma\,(CUF - BMD)} \qquad \text{equation(5)}$$

2) determining the water-split coefficient from pulp density data and measured water-flow rates, using the method of Lagrange to minimize differences between measured and best-fit values

3) calculating the corrected classification efficiency curve

4) determining the corrected d_{50} size using the equation for the reduced efficiency curve.

The whole procedure has been programmed on a HP-67 calculator and will eventually be available in FORTRAN.

Having standardized the procedure for calculating the model parameters, the next step is to establish working relationships between these parameters and operating conditions.

Figures 10 and 11 show the effects of the water-flow rate and throughput on the water-split coefficient and the corrected d_{50} size in pilot-plant experiments with a 10 cm hydrocyclone. The results agreed well with the behaviour predicted by the models described in the literature.(7,8) The water-split coefficient increased with increasing diameter of the spigot and decreased with increasing water-flow rate. The d_{50} size increased with increasing diameter of the vortex finder and decreased with increasing throughput.

However, a different trend was observed in practice with 37.5 cm and 50 cm cyclones. The water-split coefficient appeared to decrease with decreasing water-flow rate, and the d_{50} size appeared to increase with increasing throughput (Fig. 12 and 13). It is difficult to explain this behaviour with the models described in the literature. There appear to be factors, not considered in the models, such as the rate of discharge of solids through the apex, that have a significant effect on the water-split coefficient and d_{50} size. Incorporating the effect of this variable in the model produced results which were physically more consistent with the known behaviour of hydrocyclones (Fig. 14).

However, the results shown in Figure 14, are too incomplete to draw any definite conclusions. The feed size also has an effect on the water-split coefficient and d_{50} size. Results of an extensive second sampling campaign are presently being evaluated to develop a much more comprehensive model of industrial hydrocyclones.

In grinding, the results appeared more conclusive. The Breakage functions determined on ores from two different mines in New Brunswick were found to be very similar, indicating that the Breakage function is not only a characteristic of an ore but also characteristic of an ore type (Fig. 15,16 and 19). Neither operating conditions nor the size of the mill used had any significant effect on the Breakage function.

The Selection function was found to be independent of the feed size. This makes the model applicable over a wide range of feed sizes and size distributions.

Varying the number of perfect mixers in the model of the Residence Time Distribution had only a small effect on the Breakage and Selection functions (Fig. 17) indicating that grinding in a continuous mill could be approximated by batch grinding. This observation lends considerable value to batch grinding experiments.

In continuous milling it is sometimes difficult to separate the effects of several variables, for example, the feed rate (FR) and the percent solids. This difficulty stems from uncertainly about the accuracy of the mean residence time (τ). In Figure 15, it would appear that both variables have an effect on the Selection function. Whatever the mean residence time, the product of the Selection function and the mean residence time are constant for a given set of data. When the results of Fig. 15 were plotted as the product $S\tau$, in Fig. 18, the pulp density was found to have no significant effect on the grind; the feed rate was the dominant variable. Batch grinding experiments, in which the hold-up weight (HW) and the percent solids were varied independently, confirmed this finding (Fig. 16).

It is possible that the pulp density had an effect on the Selection function in the continuous experiments as indicated in Fig. 15, but also had an equal and compensatory effect on the hold-up weight: the net results being no significant effect on the grind.

$$S \times \frac{HW}{FR} = S\tau = \text{constant} \qquad \text{equation (6)}$$

Batch grinding experiments were easier to do, simpler to control and produced the same results. In both cases the Selection function was found to decrease with increasing hold-up weight. The hold-up weight in the continuous experiments increased proportionately with increasing feed rate.

The same behaviour was observed in full-scale experiments with industrial mills (Fig. 19). The Selection function was found to decrease with increasing feed rate, as the hold-up weight increased. In the primary grinding mill, the hold-up weight appeared to be less sensitive to variations in the feed rate than in the secondary mill. Estimates of the hold-up weight in the primary mill varied from 13-15 tonnes. As a result, the differences between the Selection functions were small.

In all the grinding experiments done on this ore (laboratory, pilot plant and industrial), the product of the Selection function and hold-up weight was found to be constant, indicating that no significant improvement in grinding efficiency could be achieved by varying the operating conditions (Fig. 20) Differences in the absolute rates of breakage between the primary and secondary mills were attributed to differences in ball size and ball load. The smaller 1¼ - 1½ in. balls used in the secondary mill were more efficient, as would be expected, at grinding finer material than the 2 in. balls used in the primary mill.

However, the slopes of the curves were not changed indicating little potential for improving the grind by varying ball size.

CONCLUSIONS

Modelling has permitted systematic analysis of the results of grinding and classification in an industrial mill.

It does not appear, in the case of these ores, that grinding efficiency can be significantly improved by varying the operating conditions. However, it may be possible to change the fineness of the grind and product size distribution by altering the feed size distribution.

Classification therefore emerges as the most important factor in the grinding of these ores. The model of the hydrocyclone has not been sufficiently developed to predict optimum conditions for sharper separations. Work is in progress to improve the model.

The models of the ball mills and hydrocyclones will be integrated in a simulation algorithm to test the effects of different feed size distributions, classification efficiencies and circuit arrangements on the final grind.

ACKNOWLEDGMENTS

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Mr. G. Johnston, Concentrator Superintendent, Heath Steele Mines Ltd., for his co-operation and advice in various aspects of this research.

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Messrs. P. Lachapelle, S. Lepage, and R.B. Campbell of the Mineral Dressing Section of CANMET for their excellent technical support and dedication to the work.

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REFERENCES

1. Austin, L.G.; Luckie, P.T. and Klimpel, R.R. "Solutions of the Batch Grinding Equation Leading to Rosin-Rammler Distributions", AIME Transactions; 252: 87-94, 1972.

2. Austin, L.G. and Bhatia, V.K. "Experimental Methods for Grinding studies in Laboratory Mills"; Powder Technology; 5: 261-266, 1971/72.

3. Herbst, J.A. and Fuerstenau, D.W. "The Zero Order Production of Fines Sizes in Comminution and its Implications in Simulation"; AIME Transactions; 241: 538-548, 1968.

4. Lovenspiel, O. "Chemical Reaction Engineering" (Ch. 9 Nonideal Flow); New York, Wiley, 1972.

5. Hodouin, D.;Berube, M.A.; Marchand, J.C.; McMullen, J and Everell, D. "Modelling of Heath Steele Mines Hydrocyclones and Ball Mills",;GRAAIM, Laval University; Research Report 78-5 submitted to CANMET, 1978.

6. Klymowsky, I.B. "Balancing Grinding Circuits Using Small calculators"; CANMET, Energy, Mines and Resources Canada; Lab. Report (in preparation).

7. Lynch, A.J. and RaO, T.C. "Modelling and Scale-up of Hydrocyclone Classifiers"; XI Int. Min. Proc. Cong., Cagliari, Italy, paper 9: 1975

8. Plitt, L.R. "A Mathematical Model of the Hydrocyclone Classifier", CIM Bulletin; 114-123, Dec. 1976.

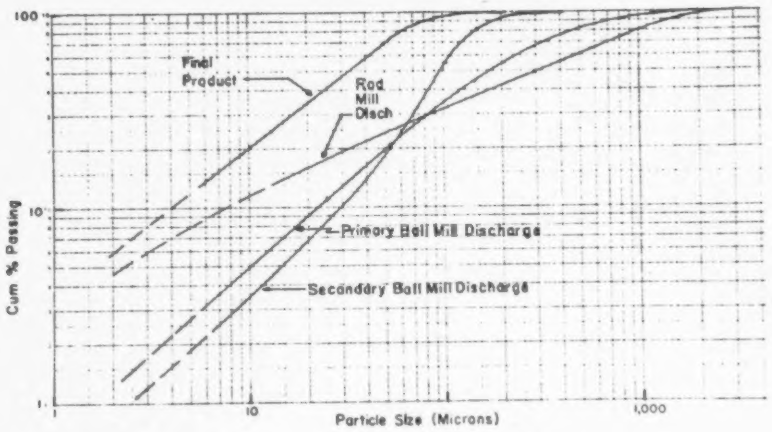

Fig. 1 Size distribution of products from an industrial grinding circuit.

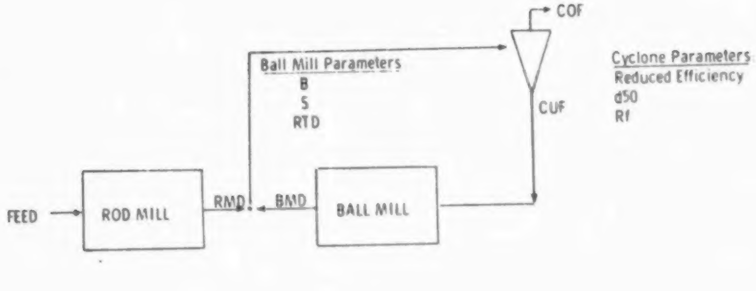
COF
Ball Mill Parameters
B
S
RTD
CUF
Cyclone Parameters
Reduced Efficiency
d50
Rf
FEED
ROD MILL
RMD
BMD
BALL MILL

Fig. 2 Grinding circuit and model parameters.

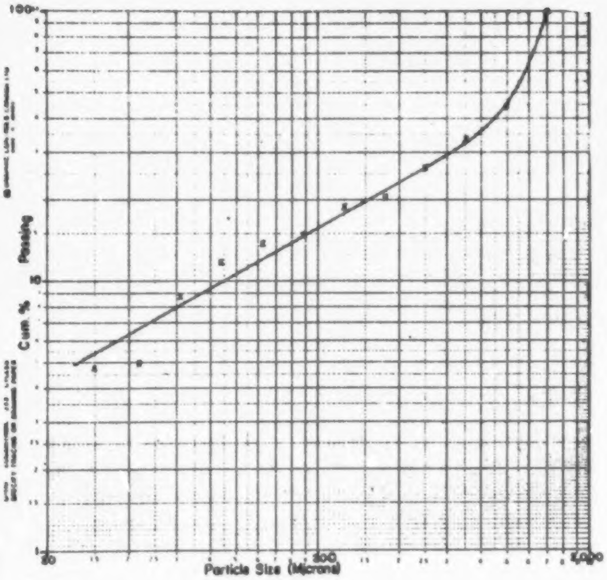

Fig. 3 Breakage function .

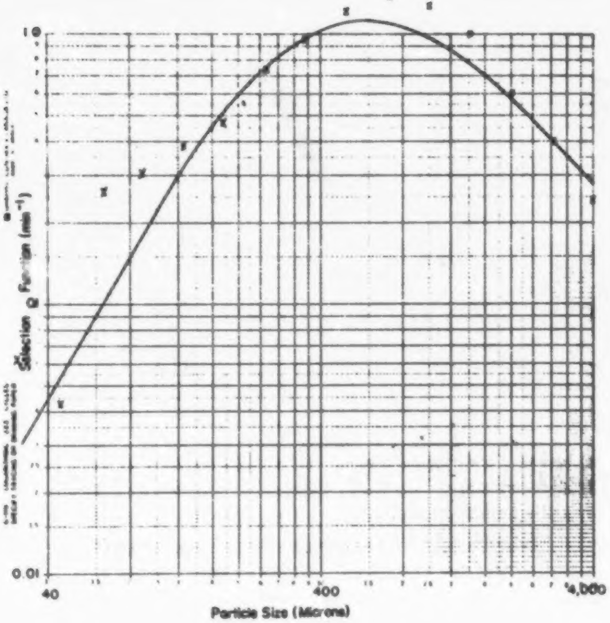

Fig. 4 Selection function.

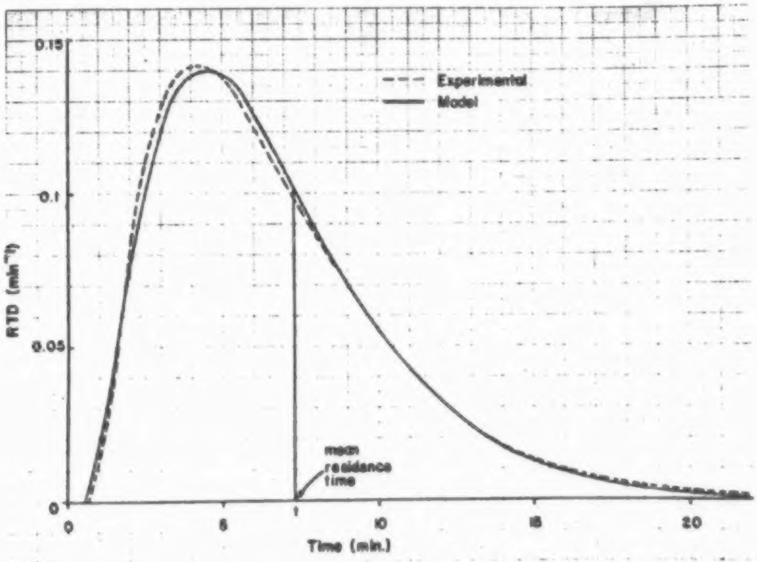

Fig. 5 Residence TimeDistribution

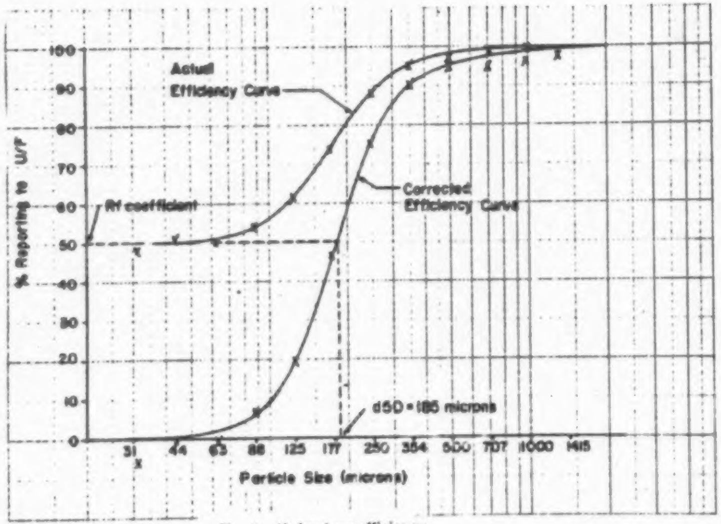

Fig. 6 Hydroclone efficiency.

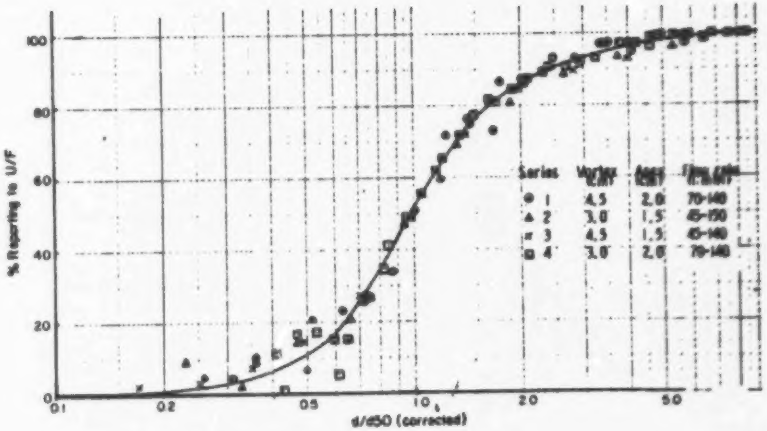

Fig. 7 Reduced-efficiency curve for a 10-cm hydroclone.

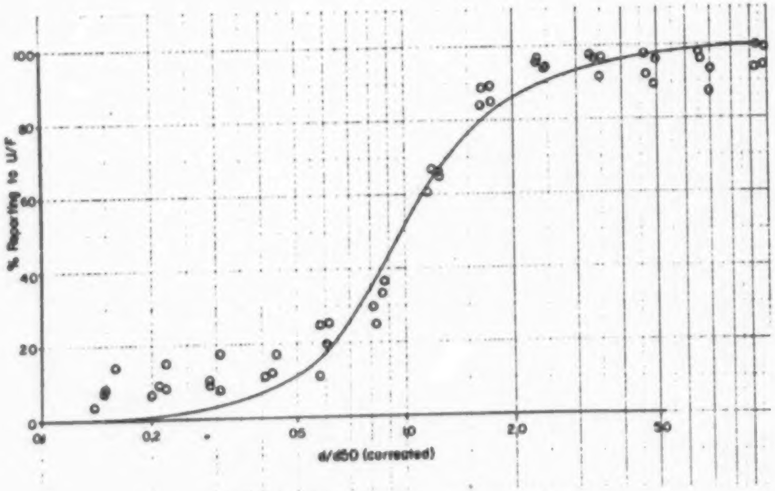

Fig. 8 Reduced-efficiency curve for a 37.5-cm hydrocyclone.

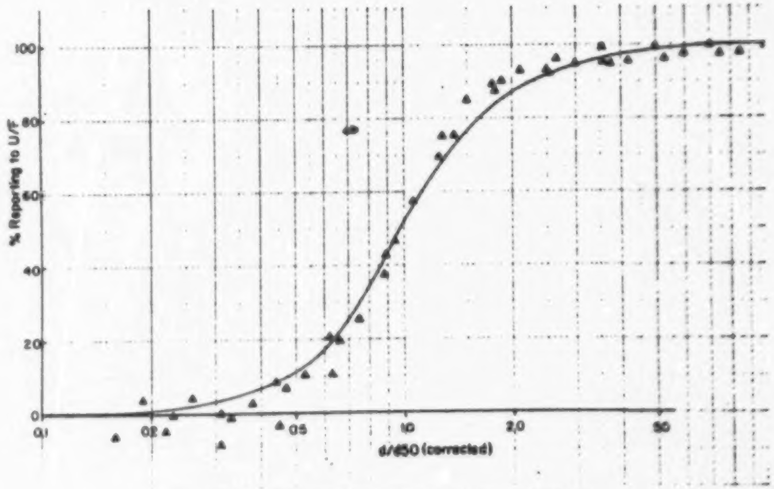

Fig. 9 Reduced-efficiency curve for a 50-cm hydrocyclone.

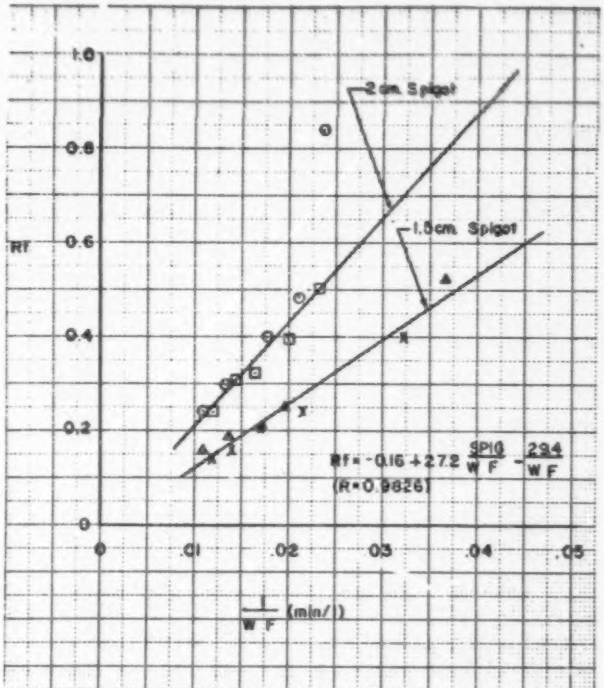

Fig. 10 Variation of the water-split coefficient with water flow rate (WF) in a 10-cm hydrocyclone.

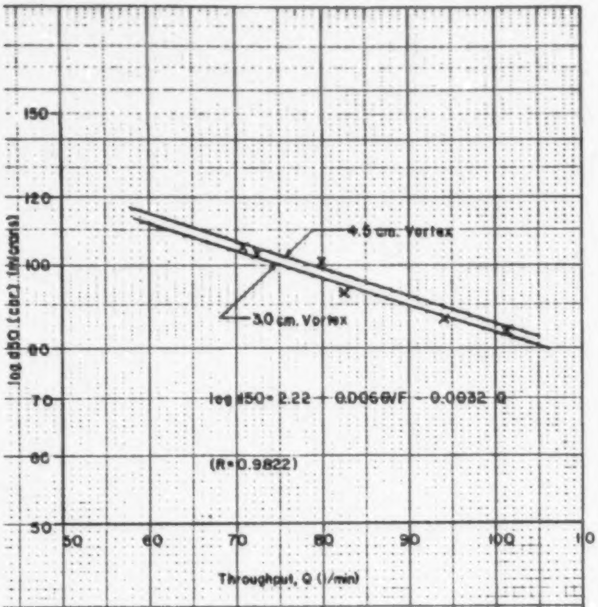

Fig. 11 Variation of the corrected d 50 size with throughput in a 10-cm. hydrocyclone.

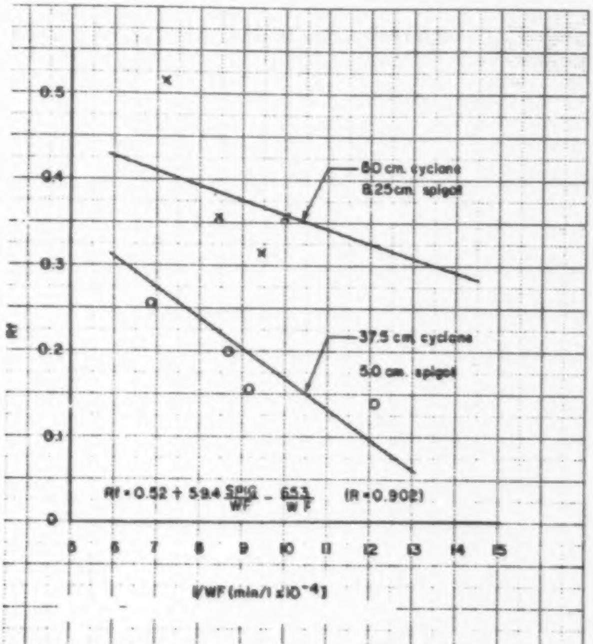

Fig. 12 Variation of the water-split coefficient with the water flow rate in 37.5 and 50 cm. hydrocyclones.

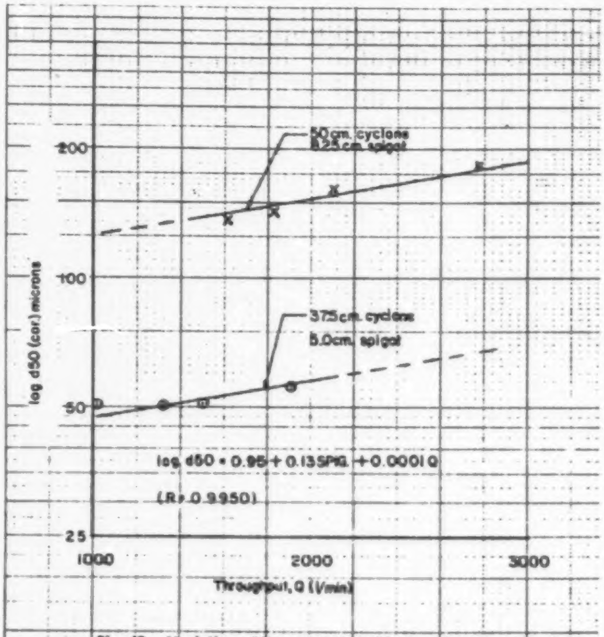

Fig. 13 Variation of the corrected d50 size with throughput in 37.5-cm and 50-cm hydrocyclones.

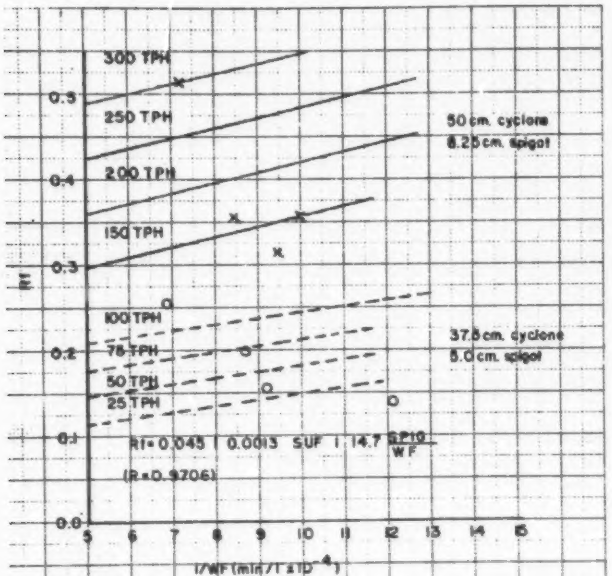

Fig. 14 Variation of the water-split coefficient with water flow rate and the solids underflow rate (SUF) in 37.5 and 50-cm hydrocyclones.

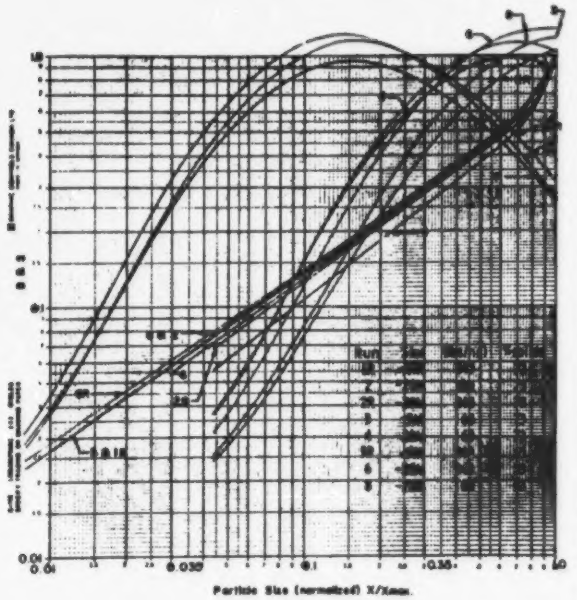

Fig. 15 Breakage and Selection functions from continuous tests in a 40-cm mill.

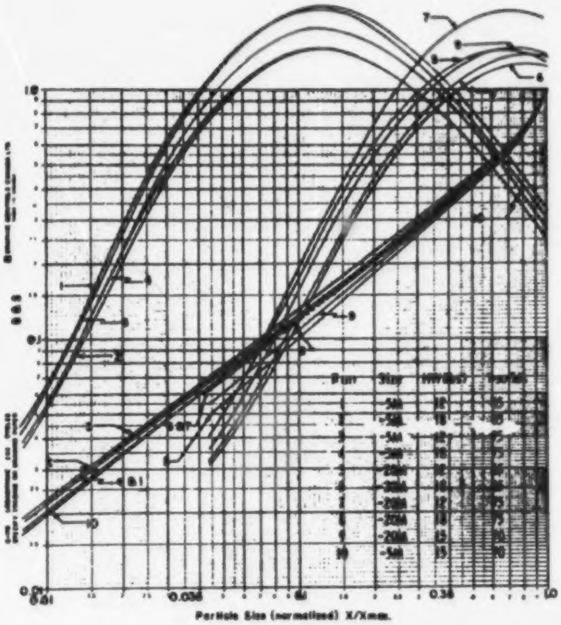

Fig. 16 Breakage and Selection functions from batch tests in a 40-cm mill.

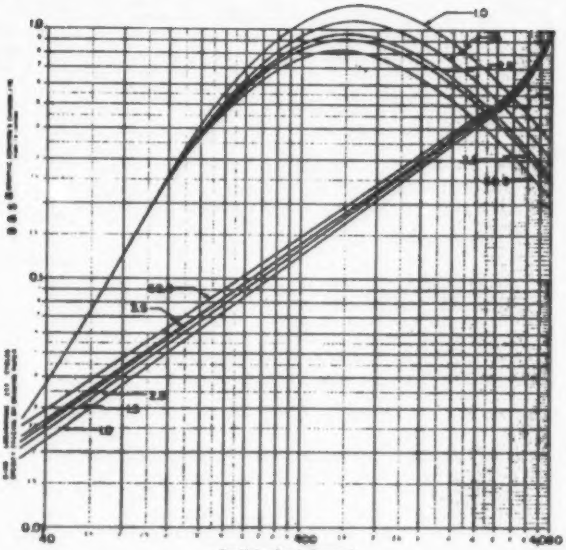

Fig. 17 Effect of number of perfect mixers on the Breakage and Selection function.

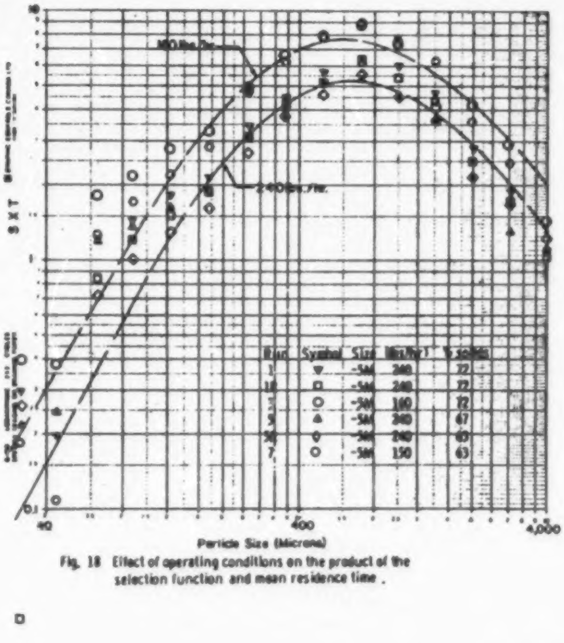

Fig. 18 Effect of operating conditions on the product of the selection function and mean residence time .

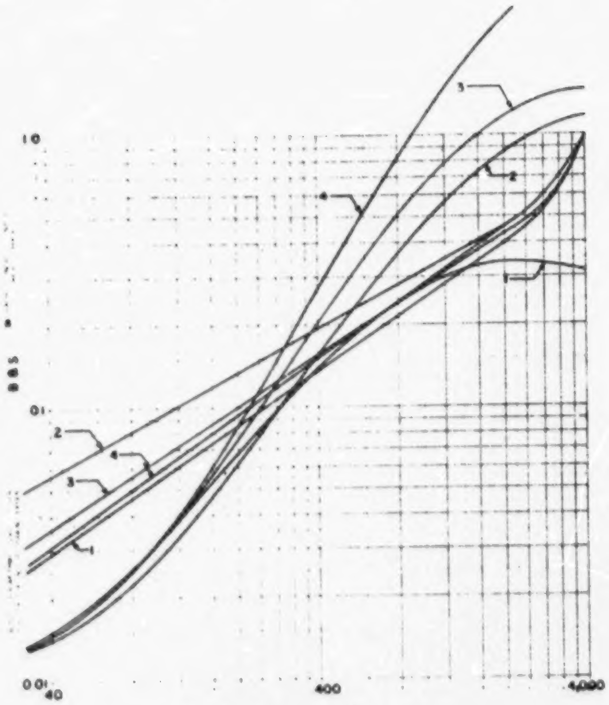

Fig. Breakage and Selection functions from industrial mill experiments.

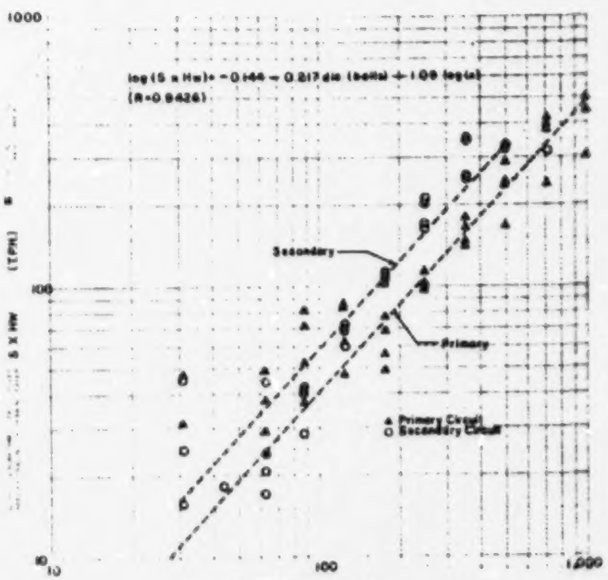

Fig. 20 Effect of operating conditions on the product of the selection function and hold-up weight.

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Paper No. 16

MATERIAL BALANCE: A STEP TOWARDS MODELLING OF FINE GRAINED SULPHIDE ORES TREATMENT

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INTRODUCTION

Although New Brunswick has the largest reserves of zinc, lead and silver in Canada, the economic exploitability of these resources has been severely hampered by the complexity of the ore bodies. Mining companies have been unable to efficiently recover high-value saleable products using existing technology. It has been demonstrated that high recoveries are possible only through the production of bulk concentrates, which, because of their complexity and relatively low grade, cannot be further processed by conventional technology. To facilitate the economic extraction and refining of these bulk concentrates it is essential that comprehensive comminution-beneficiation strategies be developed, and modeling is being used to conceive these strategies.

In the development of computer models of processing units, it is customary to calculate model parameters from experimental data describing the streams connected to the unit of interest. However, raw experimental data is rarely directly suitable for this purpose due to its inherent lack of accuracy and sometimes its unavailability. Inaccuracy of the raw data varies according to the nature of the sampling process and the location of the sampling point and would severely affect the quality of the models. Material balance calculation is used to resolve this problem by producing a set of coherent raw data. It also allows the calculation of otherwise inaccessible data. Besides the interest of the method for building up models, it is also a general purpose method that can produce metallurgical inventories.

This article aims at describing a software package and achieving two main objectives;

1) alleviating some calculation tasks which have still to be performed by hand in many processing plants and

2) conveying to the production people the progressive training needed to integrate computers as every day tools.

THE CONCEPT OF MATERIAL BALANCE

A material balance calculation has the purpose of producing a complete and accurate description of the quantitative steady state operation of an ore processing circuit, i.e. a quantitative measure of selected parameters describing the different streams of the circuit, such as chemical assay, size distribution, percent-solid, and water and solid flow-rates. Some of this information is experimentally accessible, but, especially in large streams, solid flow rates are usually only crudely approximated due to the lack of accurate on-line instrumentation. The two-or three-product formula has traditionally been used to estimate flow-rates and there is a real need for a more general and more accurate tool for calculating flow-rates: a material balance program is such a tool and operates on the same basic principle as the manual calculation as illustrated in Figure 1. This principle is to use the mass balance equations related to the unit to be balanced (i.e. the equations expressing mass conservation through the unit) to calculate unknown parameters, using experimental values to reduce the number of unknowns in these equations. An example related to Figure 1 will better illustrate this principle, using the two-product formula. Assuming that one has measured the iron percentage in the pulp of streams 1, 2 and 3 of unit I, the balance equations for this unit are:

$$\left.\begin{array}{l} w_1 - w_2 - w_3 = 0 \\ \\ Fe_1w_1 - Fe_2w_2 - Fe_3w_3 = 0 \end{array}\right)$$

therefore: $\frac{w_2}{w_3} = \frac{Fe_3 - Fe_1}{Fe_1 - Fe_2}$

Consequently, if w_1 is measured (or taken as a reference of 1 or 100) flow-rates w_2 and w_3 can be calculated with a precision which depends on the precision of the iron assays. For this reason, using calculated w_2 or w_3 to balance a second or a third unit would very quickly accumulate errors and result in meaningless values of the flow-rates. Such a hand calculation is further complicated when recirculating streams exist or when three-or four-product streams must be considered. In summary, the two-product formula allows only a rough evaluation of the flow-rates or of the flow-rates split around a unit. Modern processing plant operators need a more general method to balance simultaneously several units for periodical adjustments and general improvement of the efficiency of their operation. Several material balance programs have been designed for that purpose in the past few years. Although they may use different calculation strategies, they all perform essentially the same task:

1) calculate unknown parameters satisfying individual balance equations of each unit (also called node equations)
2) make the proper adjustment to experimental data to make it coherent with the circuit constraints and allow a solution to the above equations.

In mathematical terms, this problem is expressed as minimizing the sum of squares of adjustments made to experimental variables while observing the node equations. The Lagrange multipliers method$^{(1)}$ provides a solution to this type of problem, it states that the minimum of a function F(x), in which variable x must also satisfy equality constraints of the type $G_i(x) = 0$, is obtained for the same value of x as the minimum of the new function $F(x,\lambda) + \Sigma\lambda_i G_i(x)$, where λ_i is the Lagrange multiplier. As stated above, the function to be minimized is the sum of squares of adjustments to experimental values (e.g. $\Sigma\Delta R^2{}_{ij}$, where R_{ij} is the concentration of species j in stream i). The constraints are the node equations which usually express total mass conservation through each unit (e.g. $w_1 - w_2 - w_3 = 0$ where w_i is the solid flow rate in stream i) and conservation of individual species through each unit (e.g.: $w_1R_{ij} - w_2R_{2j} - W_3R_{3j} = 0$). The general formulation of a material balance problem can therefore be written:

minimize $F(\Delta R_{ij}, w_i, \lambda_{kj}, \lambda_k)$ (referred to as F in eq. 1)

where:

$$F = \sum_{ji}\sum WF_{ij}\ \Delta R^2{}_{ij} + \sum_{kj}\sum (Rin_{kw}\ Win_k - Rout_{kj}\ wout_k)\lambda_{kj} + \sum_k (win_k - wout_k)\lambda_k \qquad \text{Eq. 1.}$$

where

ΔR_{ij}	is the adjustment to assay R_{ij}
win_k	is the input solid flow rate to unit k
$wout_k$	is the ouput solid flow rate from unit k

Rin_{kj} and $Rout_{kj}$ are the concentrations of species j respectively in the input and output streams of unit k

λ_{kj} and λ_k are Lagrange multipliers contributing to the solution but without physical meaning (computer generated)

WF_{ij} is a weighting factor associated with assay R_{ij} and will be described later

There are several methods to calculate the minimum of objective function F(2 to 8) and we will describe only one of them which to our view, fits best the concept of the "black-box" which is expected to apply to software packages implemented in a mill for routine applications. There should indeed be such computer programs performing well defined calculations with no concern to the user about the method involved. In a similar manner, when using a pocket calculator to determine square root, there is no need to know the internal logic of the calculator, but only to follow the users manual which for instance prohibits applying the square root function to a negative number.

THE MATBAL SOLUTION TO MATERIAL BALANCE PROBLEMS

As part of its efforts to build up its resources in computer programs for simulation and modelling of mineral processes, the Ore Processing Laboratory of CANMET has acquired the FORTRAN program called MATBAL, developed by R. Wiegel at the University of Minnesota(2). Different versions of this program are being prepared and documented in a users manual in order to replace advantageously the two-or three-product formulae wherever a computer terminal is available. We will present here a typical problem with associated input and output and then pay special attention to the very often overlooked problem of confidence in the calculated values. MATBAL uses an analytical solution to the minimization of equation [1], by opposition to the direct search methods used by other programs$^{(3,5,6)}$. The solution is such that:

$\frac{\partial F}{\partial x_n} = 0$ for all variables x_n (e.g.: $x_n = \Delta R_{ij}, w_i, \lambda_k$ etc.)

The partial derivatives of F with respect to each unknown can be analytically expressed as a function of these unknowns. In MATBAL, the problem is transformed in this way into solving a system of linear equations $_A \cdot x = {}_B$

where A and B are matrices of coefficients and x is a column vector of unknowns (ΔR, w, λ). The following example illustrates a type of problem where MATBAL has proven to be robust: the calculation of mass flow rates in a flotation circuit where the streams have been assayed for one or several chemical elements. Figure 2 presents the flowsheet of the circuit to be balanced and Table 1 and 2 contain respectively the input and the output of the program. The input consists of three main subgroups of data: 1) a table identifying the streams and their assays

2) a table identifying the assays and the associated relative standard deviation (to a scale factor) and

3) a table containing the coefficients of the node equations: 1 row per equation, 1 column per stream.

These coefficients can take the values 1, 0(or blank) and - 1 depending on whether the corresponding stream is an input to, does not belong to, or is an output from the unit represented by a given node equation. For example, the second node equation is $w_2 - w_4 - w_5 = 0$ and consequently the second row of coefficients is 1, -1, -1 in columns 2, 4 and 5 and zeros or blanks in all other columns. The rest of the information consists of different codes and a description of the reference stream which in the example shown, is stream number 1. This stream flow rate can be fixed at 100 (or 1) if its actual value is unknown, in which case all calculated flow-rates will be relative to this arbitrary reference. The output of the program can be restricted to a table of adjusted chemical assays and the values of solid flow rates, as shown in Table 2. MATBAL can in fact perform much more complex types of material balances , but the case presented above is the most standard and can very easily be used on data from other circuits. Work is being done to further simplify the input data and perform automatically the flowsheet analysis to obtain the node equation coefficients. The example above was executed in 29000 Words of memory and required 30 seconds of CP time on a CDC 74 computer.

THE CONFIDENCE PROBLEM

As shown in equation [1] a mass balance calculation involves weighting factors WF_{ij} which give each assay R_{ij} a contribution to the sum of squares in proportion to its assumed or measured quality. Ideally one should use for WF_{ij} the inverse of the variance of R_{ij}, σ^2_{ij}. Unfortunately this variance is practically unmeasurable because of the magnitude of the effort involved in repetitive sampling of a full scale mill and also because of the difficulty of reproducing the same steady state conditions of operation. The weighting factors must therefore be estimated, introducing an element of subjectivity into the problem. Table 3 shows how different sets of weighting factors can affect the result, and it is usually observed that the effect is more pronounced on the calculated mass flow rates than on the adjusted assays. It is therefore highly desirable that a material balance program produce a quantitative value related to the precision of the calculated variables. Two methods have been published which develop such a feature. One, described by Smith(3) consists of an analysis of the error propagation by calculating the variance of the calculated variables corresponding to a given or measured set of weighting factors. The other, by Cutting(4) consists in studying the sensitivity of calculated variables to perturbations of the experimental data. MATBAL did not include originally such methods, but new versions will be available shortly which will calculate the precision of the output values.

Another approach to the same problem, which, although very attractive is not generally acceptable, and is mentioned here in order to point out its limitation. The calculation is started with any set of weighting factors (e.g. all weighting factors equal to one) and at the end of the calculation, the average adjustment performed by the program on, say assay Ri, is used to weight Ri in a subsequent calculation. This procedure would be repeated until there is no more variation in the calculated adjustments ΔRi. However, if we consider that the total error (identified as the total adjustment) for a given experimental value is the sum of a sampling error and an analytical error, the only error, that can be considered as characteristic of a chemical species and summed up over all measurements made of this species, is the analytical error. The sampling error is generally independent of the species, but specific to a couple "species, stream". It is therefore only when the analytical error is dominant that we can average the relative adjustments over all streams for a given assay and use this average as a measure of the quality of the experimental value to assign a new weighting factor to the corresponding data. In this special case, the proper weighting factors can also be experimentally determined by testing the analytical method of interest. Most of the time the sampling error is dominant and this iterative method will not converge. We have successfully applied this procedure to some data and observed a final convergence towards a set of weighting factors which was independent of the starting set and closely related to the weighting factors calculated from the variance of the analytical methods involved. This behaviour should not be expected in any general case.

CONCLUSION

It has been our purpose in this article devoted to material balances, to present computer methods as a powerful alternative to traditional balancing of individual units by hand calculation whether refined data for computer modelling or routine metallurgical inventory is needed. Although some circuits may present specific problems for the method, some black-box computer programs are already available to process standard cases such as flotation circuits. These programs are constantly upgraded to improve their robustness, simplify their operation and include new techniques of evaluating results. We are indebted to R. Wiegel for providing the original version of MATBAL, R. Pilgrim for programming assistance, W. Petruk for data used as example, I.B. Klymowsky and the group GRAAIM of Laval University for useful discussions.

REFERENCES

1. Saaly, T.L. and Bram, J. "Non linear mathematics"; McGraw Hill 1964- , p. 93.

2. Wiegel, R.L. "Advances in mineral processing material balances"; Can. Met. Quart; 11-2-72; p. 413.

3. Smith, H.W. and Ichiyen, N. "Computer adjustment of metallurgical balances"; C.I.M.M. Bulletin; Sept. 1973; p. 97.

4. Cutting, G.W. "Estimation of interlocking mass-balances on complex mineral beneficiation plants"; Int. J. Min. Process. 3, 1976; p. 207.

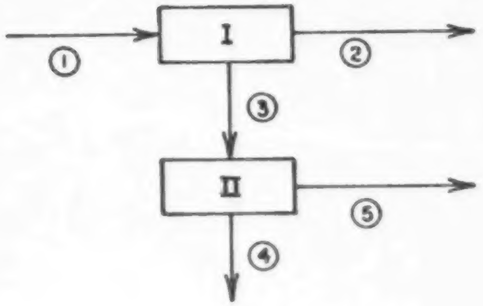
5. Mular A.L. et. al. "Mass balance of a grinding circuit"; C.I.M.M. Bulletin; Dec. 1976, p. 124.

6. White J.W. et. al. "A useful technique for metallurgical mass balances applications in grinding"; Int. J. Miner. Process. 4; 1977; p. 39.

7. Hockings, W.A., Callen, R.W. "Computer program for calculating mass flow balances of continuous process streams"; Paper pres. at the 1977 SME fall meeting and exhibit, St. Louis; Oct. 1977.

8. Tipman, R., Burnett, T.C. "Mass balances in mill metallurgical operations"; paper pres. at the 1978 CMP meeting, Ottawa; Jan. 1978.

TWO-PRODUCT FORMULA



(FOR UNIT I)

- MEASURE DATA R1, R2, R3, W1

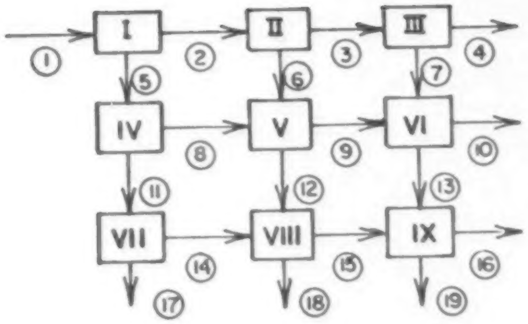
- OBSERVE CONSTRAINTS

  $W1 = W2 + W3$
  $W1R1 = W2R2 + W3R3$

- CALCULATE UNKNOWNS W2,W3

(Ri: ASSAY RELATED TO STREAM i
Wi: SOLID FLOW-RATE IN STREAM i)

MATERIAL BALANCE



(FOR UNITS I TO IX SIMULTANEOUSLY)

- MEASURE DATA R_{ij}, $i = 1$ to 19
  $j = 1$ to .. (option)
- OBSERVE CONSTRAINTS

  $W1 = W2 + W5$; $W2 = W3 + W6$ .... etc.

  $W1R_{1j} = W2R_{2j} + W5R_{5j}$ $j = 1$ to ..(option)

  $W2R_{2j} = W3R_{3j} + W6R_{6j}$

  ....etc
- CALCULATE UNKNOWNS (e.g. W_i, $i=2$ to 19) AND MININIZE DATA ADJUSTMENT: SUM OF $\Delta R^2{}_{ij}$ = minimum

(R_{ij}: ASSAY RELATED TO SPECIES j IN STREAM i)
(W_i: SOLID FLOW-RATE IN STREAM i)

Figure 1. TWO-PRODUCT FORMULA AND MATERIAL BALANCE PRINCIPLE

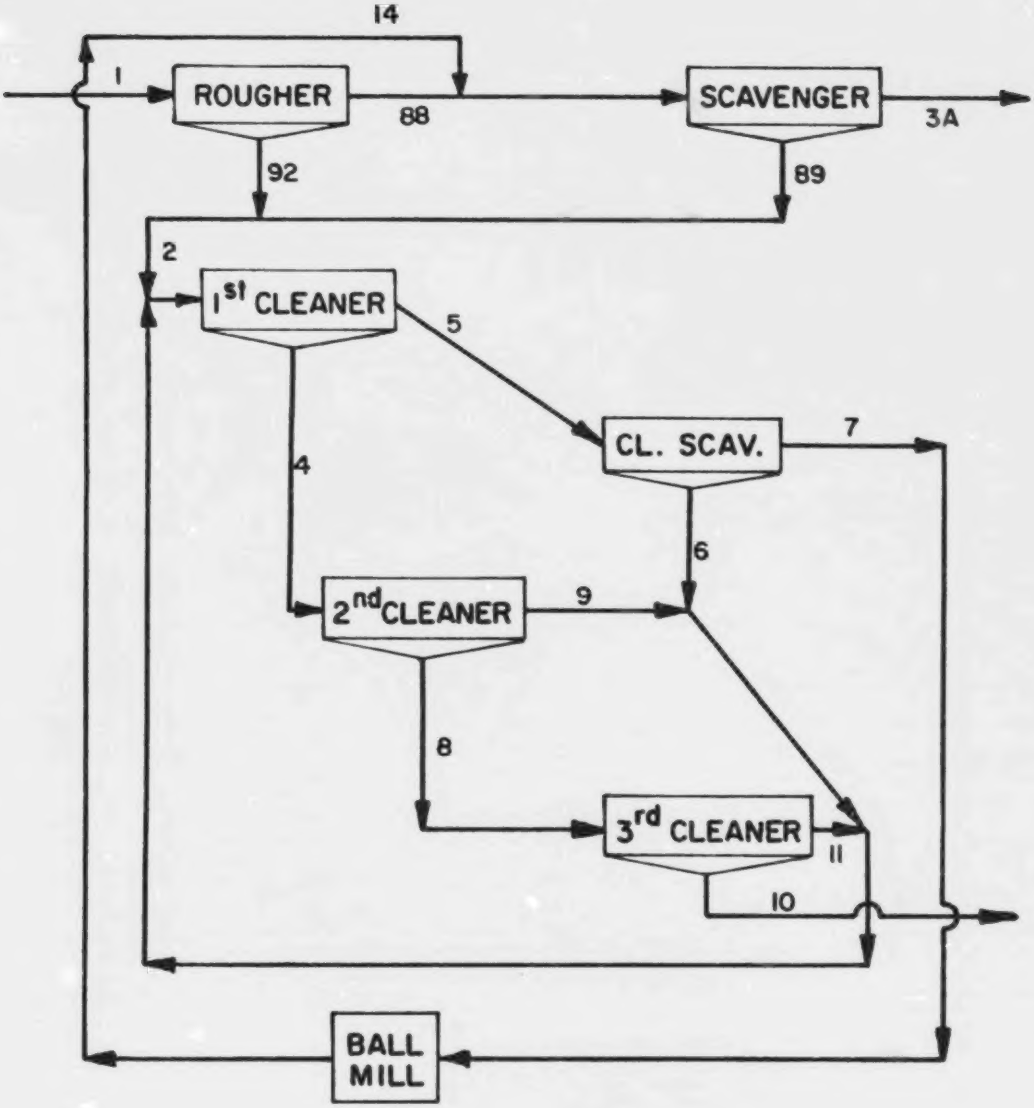

14
1
ROUGHER
88
SCAVENGER
3A
92
89
2
1st CLEANER
5
CL. SCAV.
7
4
6
2nd CLEANER
9
8
3rd CLEANER
11
10
BALL MILL

Figure 2. FLOWSHEET USED FOR MATERIAL BALANCE EXAMPLE

TABLE I. INPUT DATA FOR 15 STREAMS AND 10 ELEMENTAL ANALYSES

```
   15    7    1   10    1      0  100
 1 FEED (1)                   18  449  903 2871  105  500  700  600  980  900
 2 1ST CL CONC (4)            81 2315 1420 2358  497 1320 1000 3000 1150  700
 3 1ST CL TAIL (5)            59  949 1373 2984  303  950  700 1400 1040  600
 4 2ND CL CONC (8)           100 3487 1112 1866  628 1650  700 3100  980  500
 5 2ND CL TAIL (9)            65 1584 1464 2656  403 1300 1000 2700 1200  800
 6 3RD CL CONC (10)          110 3669 1042 1835  665 1590  600 3400  840  600
 7 3RD CL TAIL (11)           56 1729 1533 2543  393 1180 1000 2000 1140  800
 8 CL SCAV CONC (6)           90 1526 1640 2553  445 1510 1200 2400 1480  500
 9 CL SCAV TAIL (7)           49  521 1117 3394  188  830  700 1200  830  500
10 ROUGHER TAIL (88)          14  199  948 2944   59  350  600  300  850  700
11 SPLIT UP (14)              51  702 1072 3279  202  830  600 1600  840  600
12 SCAV TAIL (3A)             13  155  954 2758   50  300  600  400  780  800
13 ROUGHER CONC (92)          61 2510 1032 2548  480 1370  700 2300  970  600
14 SCAV CONC (89)             73 1089 1331 2986  310 1140  800 1900 1220  700
15 RO + SCAV CONC (2)         67 1688 1207 3014  393 1280  800 2000 1200  600
 1 COPPER                          14
 2 LEAD                            10
 3 ZINC                            16
 4 IRON                             9
 5 SILVER       PPM/100             7
 6 ANTIMONY     PPM/100            13
 7 INDIUM       PPM/10            100
 8 BISMUTH      PPM/10             35
 9 TIN          PPM/100            19
10 MERCURY      PPM               125
 1                    -1       -1                   1
 2          1    -1 -1
 3               1    -1 -1
 4             1          -1 -1
 5       1                      -1       -1
 6       1                         1 -1          -1
 7                    -1       -1          1  1
 1  100     .00001 1
 1
```

Chemical assays for:
Cu,Pb,Zn,Fe,Ag,Sb,In,
Bi,Sn,Hg, in 15 streams

Identification of assays
and relative standard
deviation

Stream Coefficients
defining Node equations

TABLE II

TABLE OF ADJUSTED ASSAYS AND CALCULATED SOLID FLOWRATES

M	STREAM (M) ASSAY:	1	2	3	4	5	6	7	8	9	10	W
1	FEED (1)	.19	4.54	9.46	28.14	1.05	4.51	6.35	5.47	8.67	7.62	100.00
2	1ST CL CONC (4)	.79	22.90	13.59	23.63	4.96	14.07	9.16	29.26	11.25	7.10	28.80
3	1ST CL TAIL (5)	.62	9.29	13.71	30.86	2.96	10.66	8.45	15.65	11.39	5.56	25.43
4	2ND CL CONC (8)	1.00	34.08	11.27	19.00	6.43	16.60	6.98	32.30	9.79	5.79	11.08
5	2ND CL TAIL (9)	.66	15.91	15.04	26.52	4.04	12.48	10.52	27.37	12.17	7.92	17.72
6	3RD CL CONC (10)	1.04	35.89	10.83	18.32	6.70	17.12	6.64	33.63	9.61	5.58	10.00
7	3RD CL TAIL (11)	.56	17.31	15.33	25.37	3.92	11.77	10.04	19.96	11.41	7.80	1.08
8	CL SCAV CONC (6)	.87	15.47	16.41	25.23	4.51	13.93	10.30	22.06	14.00	5.12	10.14
9	CL SCAV TAIL (7)	.45	5.19	11.91	34.60	1.92	8.49	7.22	11.40	9.65	5.85	15.29
10	ROUGHER TAIL (88)	.14	1.96	9.39	28.45	.59	3.45	6.32	3.25	8.66	7.88	89.01
11	SPLIT UP (14)	.53	6.97	10.81	32.40	2.02	7.99	5.98	16.80	8.28	6.02	18.69
12	SCAV TAIL (3A)	.12	1.56	9.18	28.98	.50	3.20	6.11	3.75	8.33	7.81	93.40
13	ROUGHER CONC (92)	.62	25.42	9.99	25.56	4.77	13.12	6.62	23.40	8.79	5.48	10.99
14	SCAV CONC (89)	.74	11.11	12.63	30.16	3.07	10.97	7.27	17.73	10.29	5.95	14.29
15	RO + SCAV CONC (2)	.69	17.33	11.49	28.16	3.81	11.91	6.99	20.19	9.64	5.74	25.28

TABLE III. EXAMPLE OF THE EFFECT OF DIFFERENT WEIGHTING FACTORS ON THE RESULT OF A MATERIAL BALANCE CALCULATION

Stream	Cu%	Pb%	W	Cu	Pb	W	Cu	Pb	W
1.	.19	4.52	100.	.19	4.51	100.	.19	4.68	100.
2.	.79	23.06	28.14	.80	23.11	30.03	.78	22.46	32.87
3.	.63	9.44	26.82	.62	9.19	28.69	.59	8.41	23.42
4.	1.00	34.14	11.08	1.00	33.54	12.32	1.01	34.90	11.17
5.	.66	15.87	17.06	.66	15.85	17.17	.66	16.05	21.70
6.	1.05	36.12	9.91	1.06	35.90	10.75	1.03	35.56	10.76
7.	.56	17.31	1.17	.56	17.33	1.57	.56	17.29	.40
8.	.86	15.32	11.27	.87	15.56	11.07	.90	16.05	6.63
9.	.45	5.18	15.55	.46	5.20	17.62	.47	5.39	16.79
10.	.14	1.96	89.08	.14	1.96	89.41	.14	1.93	88.06
11.	.53	6.96	19.10	.52	6.94	25.78	.52	6.92	22.30
12.	.13	1.56	93.65	.13	1.57	97.41	.13	1.57	94.74
13.	.61	25.38	10.92	.61	26.04	10.59	.61	24.99	11.94
14.	.74	11.09	14.53	.74	11.35	17.78	.75	11.20	15.62
15.	.69	17.22	25.46	.69	16.83	28.37	.69	17.17	27.56
	σcu = 14 σPb = 6			σcu = 6 σPb = 14			σcu = 1 σPb = 1		

w = calculated solid flowrate

σ = relative standard deviation (%)

Cu, Pb: adjusted assays

Discussion of paper by I. Klymowsky and D. Laguitton

G. Rodger: What was the reason for choosing the Wiegel material balance program over others ?

Reply: It is a very rigid analytical program that is simple to use if you use it in the range for which it was designed. Otherwise a considerable rewrite is necessary. Other programs are more difficult to use but are more versatile.

Paper No. 17

CHEMICAL ADDITIVES FOR IMPROVING THE WET GRINDING OF ORES

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WHITEHORSE COPPER MINE
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Presented at 1979 Annual Meeting of the Canadian Mineral Processors
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Introduction

Laboratory and industrial grinding tests have shown that the process of size reduction of metal containing ores can be significantly influenced by chemicals added to the slurry being ground. The terms grinding aid or grinding additive refer to a substance which, when mixed into the mill contents, causes an increase in the rate of size reduction. The increased rate can be used to grind a higher feed rate to the desired product size or it can be used to produce a finer product size at a fixed rate. Whether the use of a grinding aid is justified in any given situation depends on the cost of the substance versus the improvement of output or product quality obtained with its use.

A wide variety of grinding additive mechanisms have been suggested in the general mining literature (1,2,3,4,5,6,7). The prevention of particle agglomeration, slurry cushioning of grinding media impact, grinding media coating, changing the strength of macro porous rocks, and chemical adsorption on flaw surfaces to change bonding forces and surface energy have all been suggested. More recently, some grinding additive work has been carried out (8,9,10,11) which suggests that the use of efficient chemicals to control the fluidity of dense slurries is a viable economic approach to achieve improved grinding rates. These increased grinding rates are achieved with no increase in power to the grinding mills and apparently with no increase in media consumption costs. Unlike much of the previous work with grinding additives, this recent work is accompanied by a rather complete engineering characterization and prediction of performance capability. The chemicals involved were chosen so as to efficiently control the viscosities of slurries under conditions of high shear yet not adversely affecting flotation, flocculation, and other downstream surface chemistry operations.

In particular, this paper will concentrate on results achieved in copper and taconite ore grinding using an experimental grinding aid product of The Dow Chemical Company designated as XFS-4272. The major tests of XFS-4272 to date have occurred in Chile, Arizona, Minnesota, and recently in Canada at Whitehorse Copper Mines. Examples of lab testing and industrial scale trials will be discussed so as to fully demonstrate the type of results that have been achieved.

The grinding circuit at Whitehorse Copper Mines was selected as a test site for the fluidity control approach because of a number of favorable conditions: a recent ore matrix change leading to a highly viscous slurry (hence a significant dropping off of throughput to maintain grind and recovery), the need to lower % solids to maintain ball mill stability and hydrocyclone efficiency, and a single plant grinding circuit configuration with dedicated flotation and water recovery systems. The grinding circuit is illustrated in Figure 1. Thus at Whitehorse Copper, all the factors, which should be affected by a fluidity control approach, were already existing in their normal operation. Generally, in industrial testing to date, it has been necessary to deliberately move the circuit into a region of high viscosity in order to demonstrate the advantages of fluidity control grinding aids over normal but less viscous operations.

Also, the availability of dedicated flotation and water recovery systems is useful as undesirable side effects of using XFS-4272 could be more easily observed than in many plants.

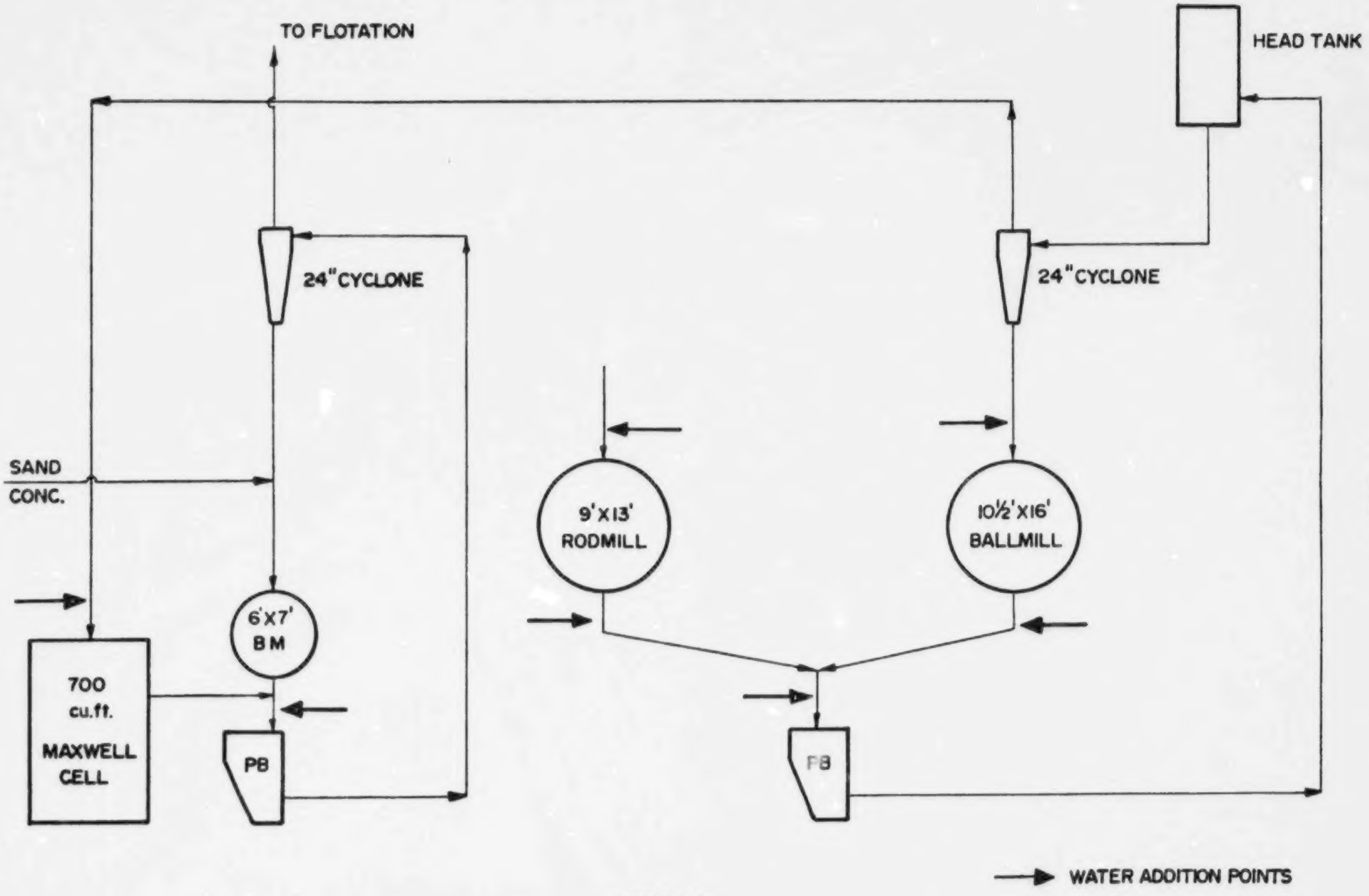

TO FLOTATION
HEAD TANK
24"CYCLONE
24"CYCLONE
SAND
CONC.
9'X13'
RODMILL
10½'X16'
BALLMILL
6'X7'
BM
700
cu.ft.
MAXWELL
CELL
PB
PB
WATER ADDITION POINTS

FIGURE I

WHITEHORSE COPPER MINES LTD.

GRINDING CIRCUIT

Laboratory Scale Wet Grinding of Ore with Water Only

The most important grinding additive is, of course, water. It is well known, in practice, that grinding in water is advantageous over dry grinding in tumbling ball mills. Industrial experience has shown that the modification of slurry density by water addition or removal is the easiest and most rapid technique for influencing slurry transport in grinding mills. As part of the slurry fluidity studies mentioned earlier, the effect of slurry density modification with water only on the breakage processes in tumbling media mills was studied. Such a characterization is necessary to give a realistic basis for the explanation of chemical additives.

In brief, it was found (8,9,10,11) that by varying the water to solids ratio in a mill of fixed slurry volume (e.g. a batch or a continuous overflow discharge mill), the rate of breakage of different sizes changed in a uniform manner but the daughter fragment distributions of breakage did not change appreciably. To assist in data analysis, the well-known concepts of specific rate of breakage of size i material, S_i, and cumulative daughter fragment distributions of size j material broken to less than size i material, B_{ij} were employed (12,13). Table 1 gives a typical laboratory scale result of this work. The values of B_{ij} (the distribution of daughter fragments from breakage) were unaffected by slurry density in the range indicated, but the fractional rate of breakage, for example, S_1 for the largest size interval, consistently decreased with increasing % solids. However, in the normal range of operation, this decrease in S was compensated by an increase in the weight of charge W so

Table 1. Comparison of SW values for wet grinding of a copper/molybdenum sulfide ore in a ball mill of 8-inch diameter (fraction ball filling = 0.33, 1 in. steel balls, mill diameter = 200 mm, speed of mill = 67% critical). S_1 is the fractional rate of breakage of the largest sieve size fraction of feed material and W is the weight of ore charge in the mill.

Specific Gravity of Ore	Mill Volume	Slurry Density		S_1	W	S_1W
	cm^3	wt% solids	vol% solids	min^{-1}	g	g/min
2.60	5790	60	36.1	0.433	1000	433
Formal bulk density = (2.60)(0.6) = 1.56 g/cm^3		62	38.1	0.417	1054	440
		64	40.1	0.398	1110	442
		68	44.5	0.351	1231	432
		72	49.2	0.297	1362	405
		74	51.8	0.272	1434	390

Table 2. Effect of fluidity modifier on absolute rates of breakage S_1W for taconite ore in a laboratory ball mill (see Table 1 for conditions).

Specific Gravity of Ore	Concentration of Additive	Slurry Density		S_1	S_1W
	g/kg solid	wt% solid	vol% solid	min^{-1}	g/min
3.30	0	80	54.9	0.327	616
	0	82	58.1	0.305	608
	0	84	61.5	0.270	570
	0.2	84	61.5	0.290	612
	0.6	84	61.5	0.332	700

that S•W is almost a constant. It isn't until one reaches a relatively high % solids, e.g. 72%, that the production rate, e.g. S_1•W, starts to fall.

Figure II illustrates a typical pattern in the shift of the breakage curves with changes in slurry density. Note the S curves are parallel indicating the breakage mechanism is the same in all cases, only the absolute rate differs. When the net production of material less than some size is measured (e.g. 44μ), the value is again almost constant for 64 and 68% but falls for 72% slurries as shown in Figure III. Going through this type of analysis is important, as the recognition of that % solids region where S•W is no longer constant, but falling, is key to the use of fluidity control grinding additives.

Laboratory Scale Wet Grinding of Ore with Fluidity Control Agents

Table 2 illustrates the importance of identifying the region where slurry viscosity has become a limiting factor. Once the value of S•W starts to fall, the addition of a grinding additive becomes possible. In fact, not only can an additive extend the % solids region over which S•W is a constant, but can actually increase the absolute value of S•W. It is this increase that represents a true effect or increase in grinding production that is possible with the type of grinding additive being discussed. Figure IV illustrates the S curves for this case but note that the increase in the S value with chemical is not taken at the expense of decreasing W as all slurries are at 84% solids. Figure V is another way of demonstrating the effects just discussed. Even though the specific example described here is for a taconite ore, exactly the same type of result is observed on copper-molybdenum sulfide ores. It is important to note that

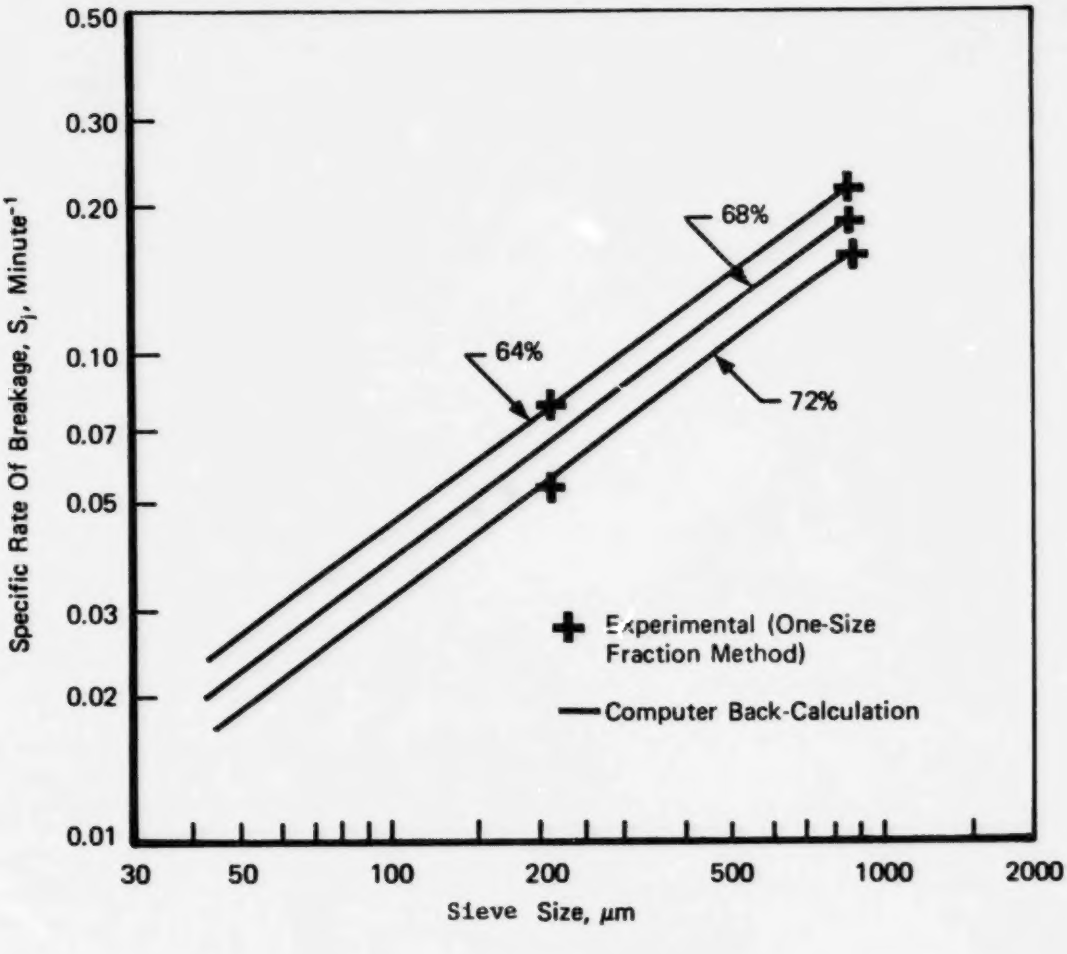


FIGURE II. Back-calculated S curves for copper ore slurries ground in 8 inch batch ball mill for various slurry densities.

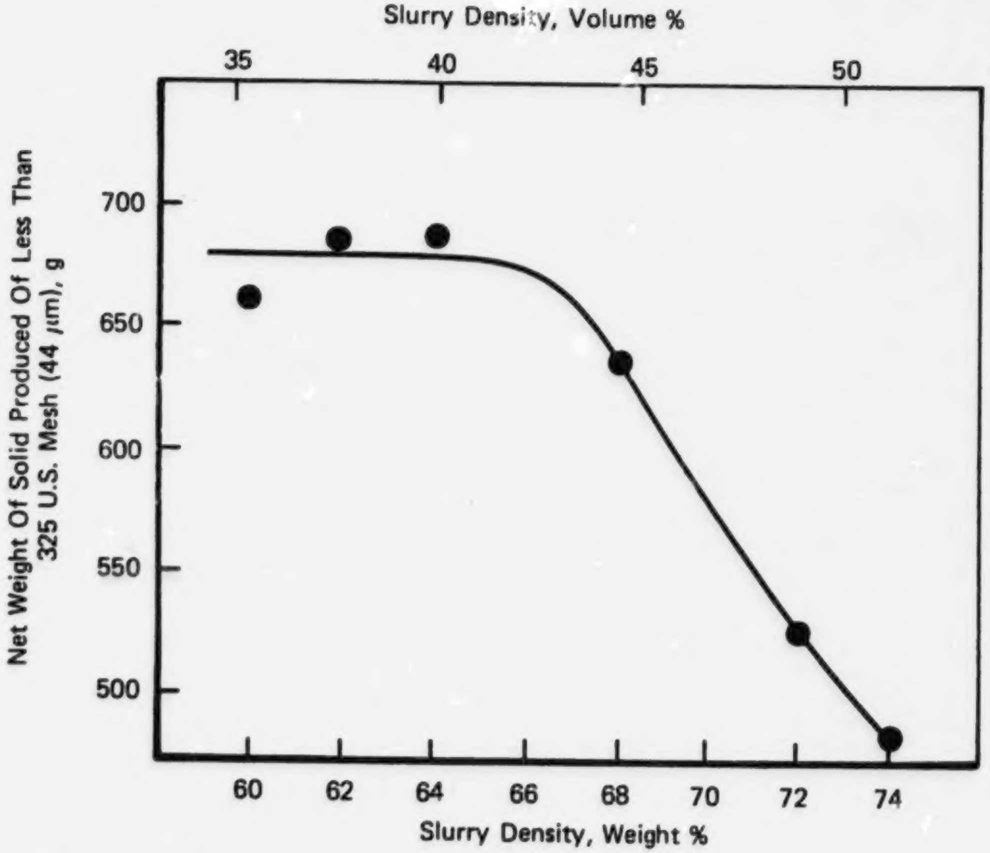


FIGURE III. Net grams of minus 44μm material obtained by grinding copper ore for 40 minutes in an 8 inch batch ball mill as a function of slurry density (constant slurry volume).

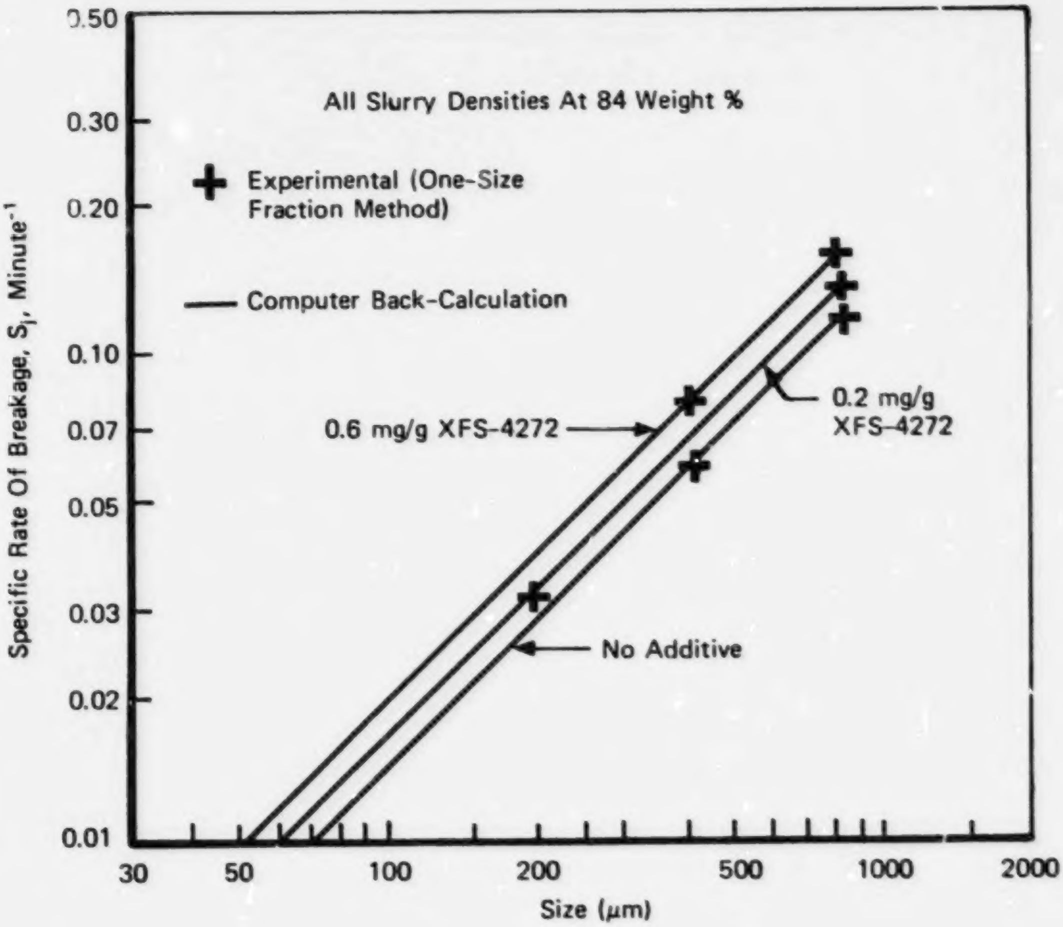


FIGURE IV. Back-calculated S curves for taconite ore slurries ground in 8-inch batch ball mill showing effect of grinding additives.

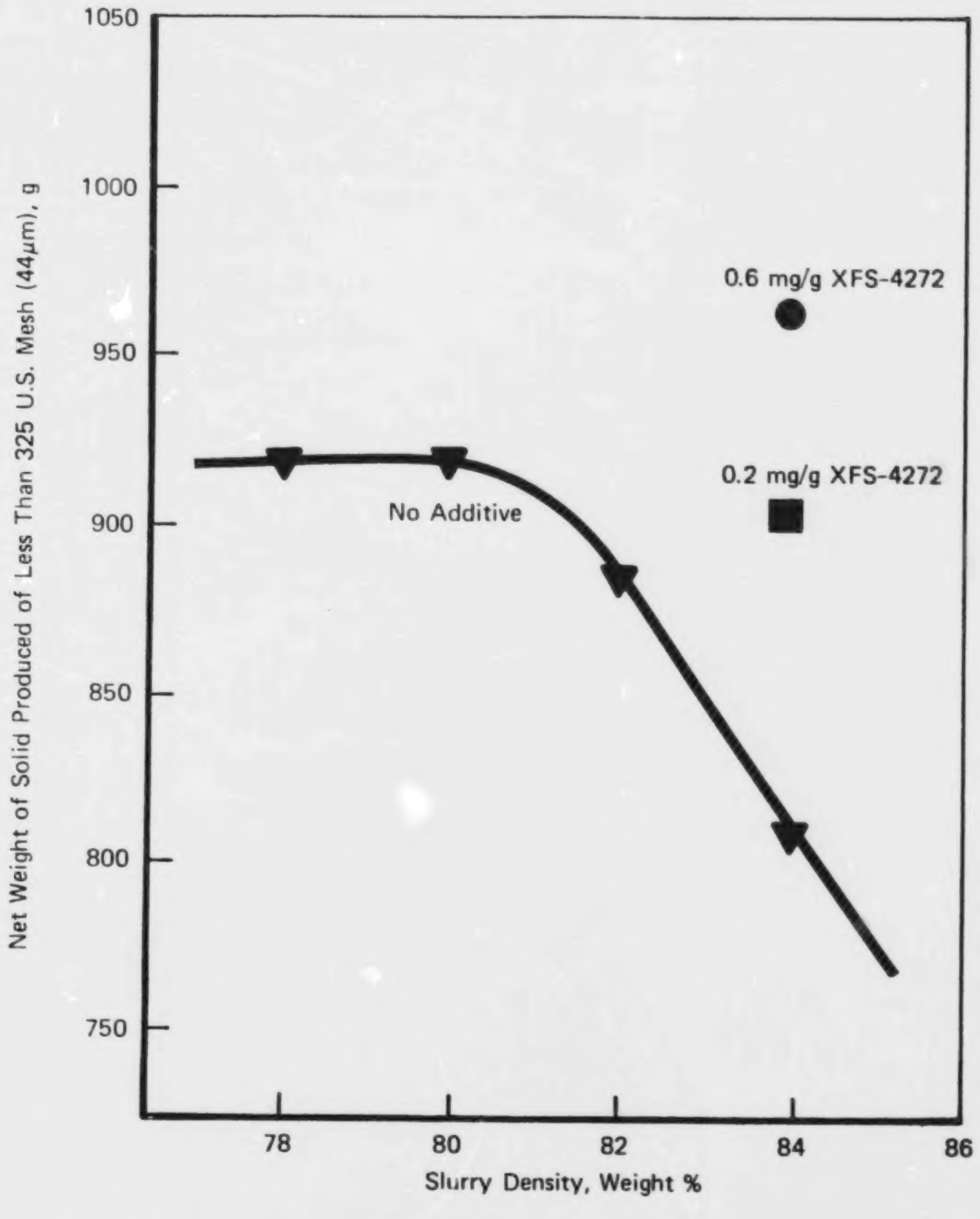


FIGURE V. Net weight of solid, in grams, less than 44μm, produced for taconite ore ground for 45 minutes in 8 inch batch ball mill.

if the % solids region is extended too far, even the use of chemical will not recover the value of S•W. The parallel nature of the S curves again demonstrates that using an additive does not change the grinding mechanism noticeably. In addition, the general slopes of the S curves, with and without the additive, are similar. In testing in the lab, the amount of chemical required in a batch mill is usually three to four times that of a continuous test because of the large surface area changes involved in batch testing (hence larger number of adsorption sites for the grinding aid).

The effect of the additive is most easily demonstrated in the lab using grinding tests involving a constant (e.g. 120% of ball void volume) slurry volume (varying wt % solids within this constant volume). It is also important to identify the effect of the slurry volume loading. Figure VI demonstrates a complete set of batch grinding experiments relating the net weight of solid produced of less than 325 U.S. mesh as a function of both slurry volume and percent solids for a fixed volume of 1 inch balls.

The data of Figure VI on taconite ore illustrates a number of trends that are typical of laboratory batch ball mill testing with fluidity modifiers like XFS-4272. At any % solids with chemical, the net weight of product produced is relatively flat and then falls off at high W loadings. With lower % solids, the net weight of minus 325 mesh material without chemical is higher than runs made at higher % solids without chemical regardless of the weight loading in the mill. Also, at low % solids, e.g. 78, 80, and 82, the net weights produced are similar indicating that S•W is constant.

At 84 and 86% solids, S•W is no longer constant and the net weights produced fall off significantly. As indicated in Figure V and Table 2, the use of XFS-4272 is effective when S•W is no longer constant. Thus at 78,

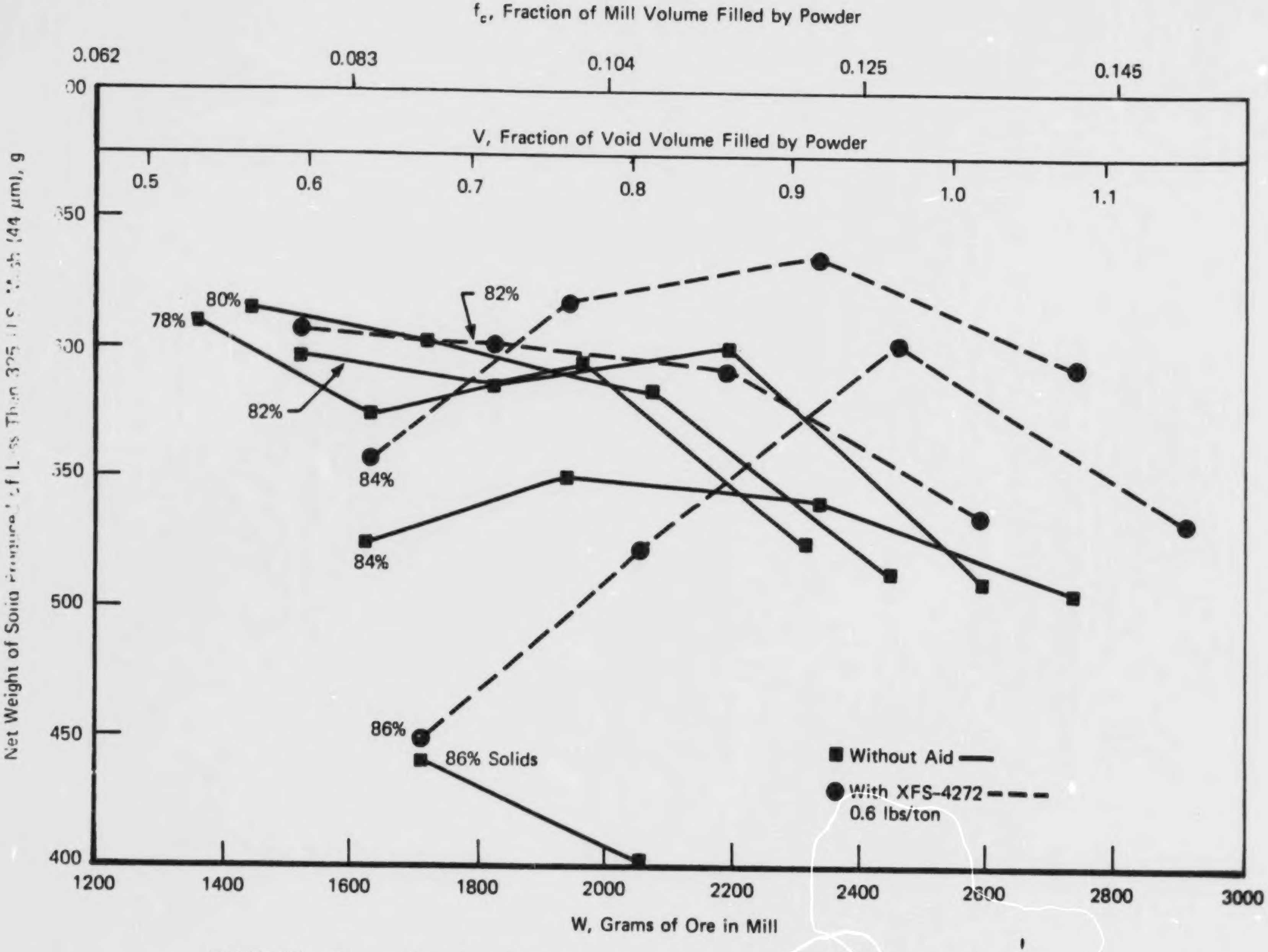


FIGURE VI. Net weight of solid produced less than 44µm for taconite ore ground for 30 minutes in 8-inch batch ball mill (14.1% <325 U.S. mesh in feed).

80, and 82% solids the net weights produced are similar with and without chemical but the runs with chemical at 84 and 86% solids do produce higher net weights of product with chemical than without.

The runs at 84 and 86% solids with chemical illustrate three important points in evaluating fluidity-control grinding additives like XFS-4272. The first point is that, in general, a relatively narrow range of % solids (Δ=2 to 4 %) region exists over which an absolute increase in net weight produced can be gained. In Figure VI, this optimum exists near 84%. At Whitehorse Copper Mine the normal operating range is already in a region of viscosity control so that the chemical will be effective with no % solids change. A second observation is that at higher % solids with chemical, an optimum ore loading in batch testing occurs which corresponds, for example, to a fraction of mill volume filled by ore of 0.10 to 0.15 for taconite. The third point is that care must be exercised not to increase the % solids too high, e.g. 86% in Figure VI. Even though the use of chemical will be beneficial under this high % solids condition, the net weight produced will not be larger than some other net weight that can be produced without chemical.

Continuous Industrial Scale Tests

Although the laboratory results suggest that increased rates of grinding can be obtained with the grinding aids, the effect was only better than water alone at very high slurry densities. Therefore, it has to be demonstrated that the fluidity modifiers also allow these high slurry densities to flow through large mills without problems. A number of industrial grinding circuit trials involving rod and ball mills with overflow discharges have been performed using these efficient chemical additives (8,9,10,11). The initial results have been very encouraging, with several trials demonstrating an ability to increase the rates of production between 9 and 15% in both open and closed circuit operations (11). This has been done while maintaining a similar product size distribution and keeping the total mill power draw the same as for no additive (hence, a lowering of energy/ton). Figure VII shows a typical result from one of these trials in the copper industry involving a ball mill in closed circuit with a classifier. The rate of make-up feed and product size distributions were monitored continuously. All other factors such as percent solids, classifier control, etc. were kept constant over the duration of the test.

Several important conclusions can be drawn from inspection of Figure VII. The first is the ability of the additive to cause an increase in fineness of grind with approximately constant feed rate; the period from $\hat{2}$ to $\hat{3}$ produced an average size distribution which had 4% finer than the desired control size as compared to the period $\hat{1}$ to $\hat{2}$. When the additive was shut off, the % greater than the control size went back to its original level. At time $\hat{4}$, the additive was again used with a subsequent increase in fineness of grind. After time $\hat{5}$, the feed rate was slowly increased. After 3 hours with the aid, the feed rate was 8% higher than earlier, but with the product size being the same as the base case without additive. When the additive was shut off and the feed rate reduced, the mill output returned to approximately its earlier operating condition. Also, during the period of increased feed rate with the grinding additive, the mill power draw was identical to the base case without chemical. This has been shown to be true (11) for all the overflow ball and rod mills investigated

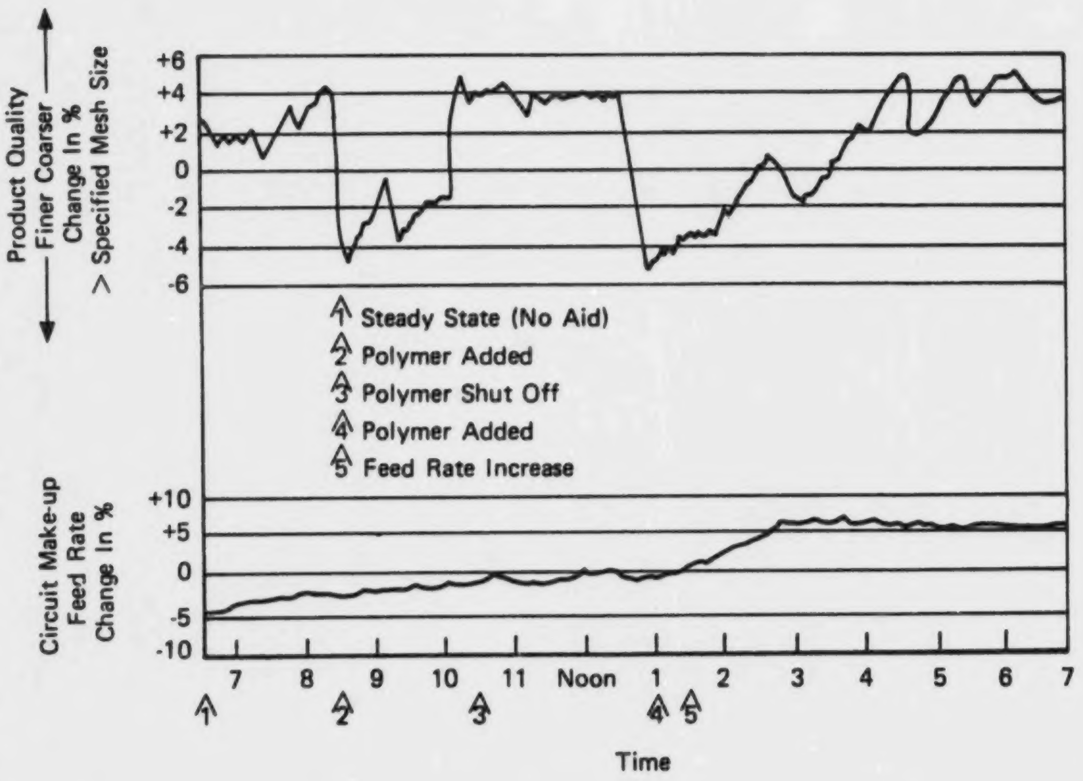


FIGURE VII. Typical response curve for industrial scale mill.

where the mass of grinding media and slurry is constant during operation. Also, it appears from preliminary data that no increase in media consumption is experienced with the use of XFS-4272.

Figure VIII shows the net production rate of minus 35 mesh material in the open circuit continuous rod mill of Whitehorse Copper. The water addition was held constant so that Figure VIII represents essentially the continuous counterpart of the batch lab data of Figure V. The optimal operating conditions without chemical was between 95 and 100 tph at the water level chosen (80 gpm). The grinding additive performance is rather obvious at all feed rate values tested.

Figure IX shows a typical output of the total Whitehorse Copper circuit, with and without XFS-4272, with the water being held constant. Obviously, the ore hardness varied widely during the test and the chemical was evenly split between the rod and ball mills with a total dosage of about 0.5 lbs/ton of feed. The increase in throughput of minus 200 mesh material with chemical is clearly evident. It is also important to note that the optimal use of chemical (amount and points of addition) had not been determined at the time this data was taken.

In routine industrial practice, a series of tests are required with rather complete circuit size distribution sampling to identify optimal water and chemical addition points. As might be expected, the amount and splitting of chemical depends on which pieces of equipment in the circuit are or become fluidity controlled (limited) by increases in throughput. This is one goal of a longer term plant trial (6 weeks, etc.). Also, the identification of flotation effects, tailings disposal data, etc. which require longer time periods to quantify need to be carried out in extended tests.

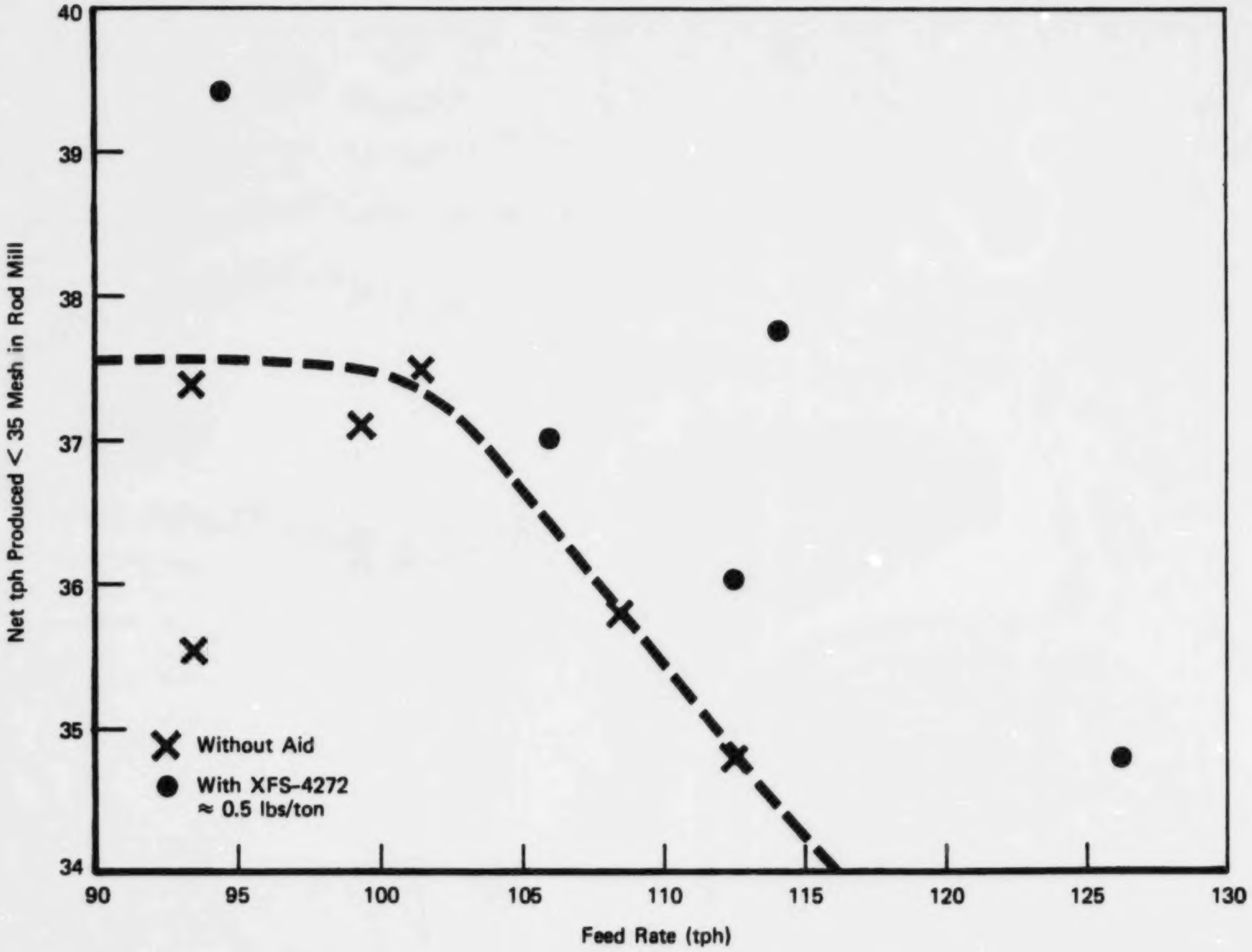


FIGURE VIII. Net production rate in tph of minus 35 mesh material in an open circuit continuous rod mill with copper ore, holding water constant.

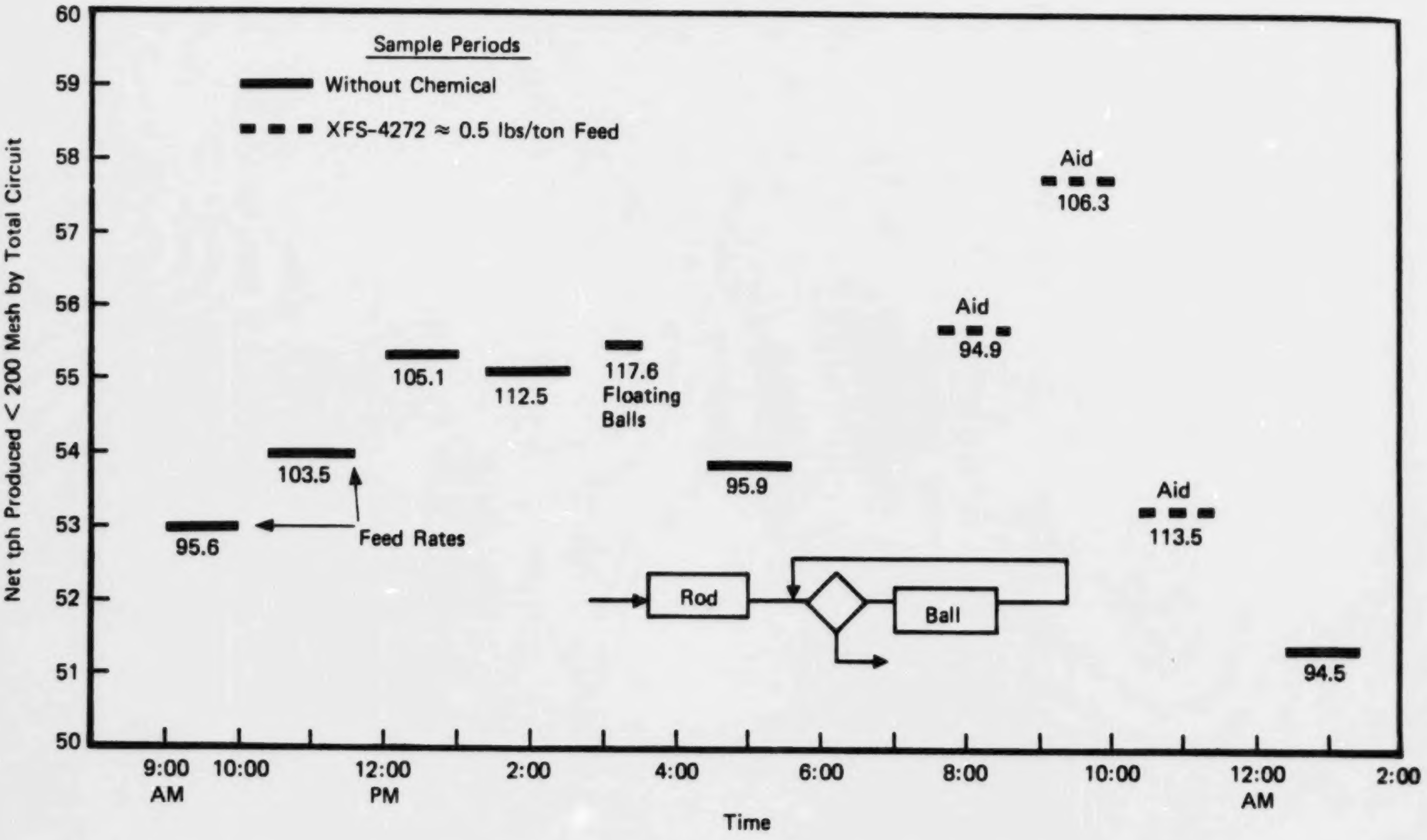


FIGURE IX. Net production rate in tph of minus 200 mesh material produced in a total circuit with copper ore, holding water constant.

The choice of whether to use grinding aids to achieve increased throughput at a constant product size or a finer product at a constant throughput, or some combination of increased throughput and fineness, depends on the economics of the particular grinding operation under study. The decision depends to a large measure on the liberation achieved at the normal grinding condition, on what finer grind could be achieved economically with chemical, and on what changes in liberation accompany the finer grind. For example, in taconite grinding, this could represent more silica liberation and removal, in copper grinding higher copper flotation recovery, etc.

Figure X shows the increased fineness of grind that was achieved with chemical in a copper sulfide ore grinding circuit at two different feed rates. This data is quite typical of the type of grind that can be achieved when all of the effect of the chemical is put into forcing finer grinding at constant throughput. At Whitehorse Copper, preliminary testing of this approach has shown an average fineness of 3% passing 200 mesh at constant feed rate to be achievable.

Figure XI gives two actual examples of the increased liberation of copper as a function of fineness of grind (as indexed by the weight % less than 200 mesh). Initial evaluations at Whitehorse Copper have shown approximately a 0.2% increase in recovery for every 1% increase in fine passing 200 mesh. This data, plus the increased % fineness produced by the additive (e.g. the data of Figure X), plus the value of the increased recovery of copper, enables a first estimate to be made of the break-even cost which can be paid for the chemical. If the chemical costs less than this, the use of additive will certainly be advantageous. If the cost of the chemical is somewhat higher than this break-even value, the use of

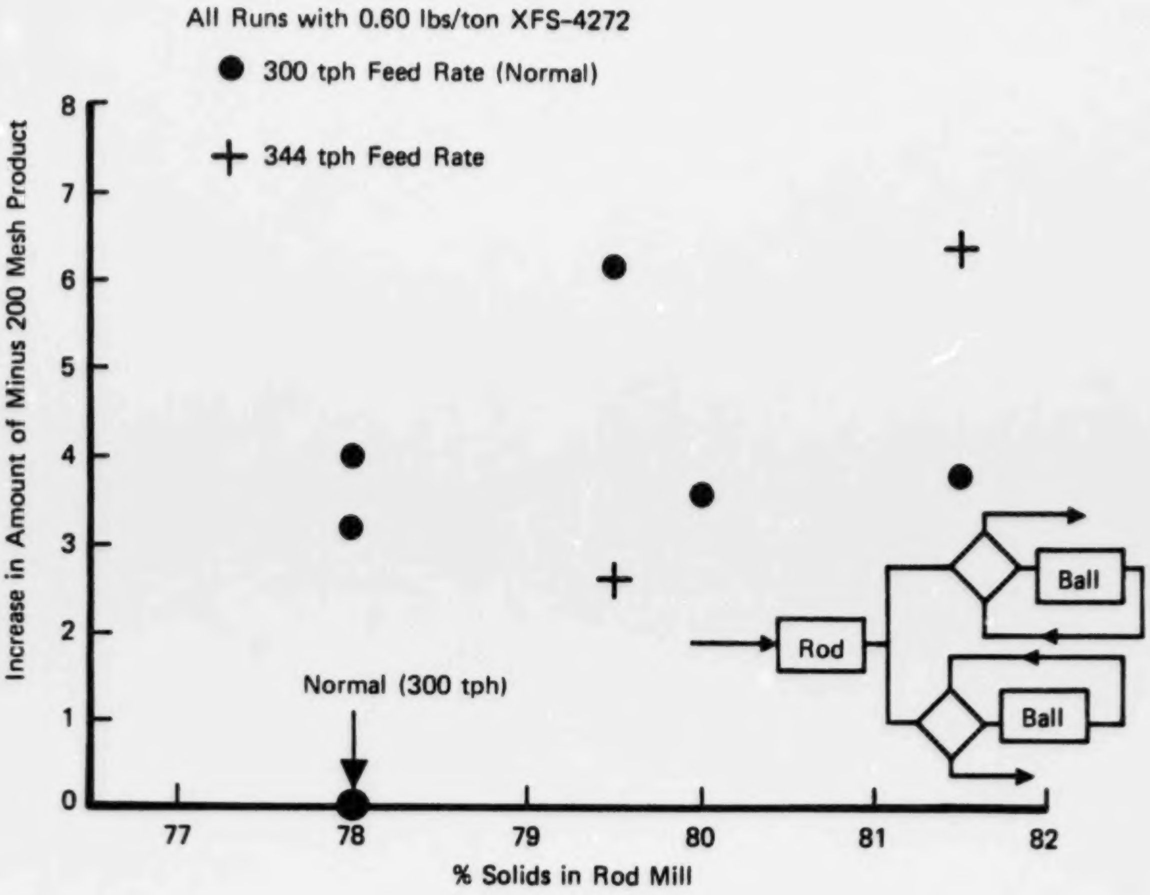


FIGURE X. Increase in % minus 200 mesh in the product to flotation from a copper sulfide grinding circuit holding feed rate constant using XFS-4272.

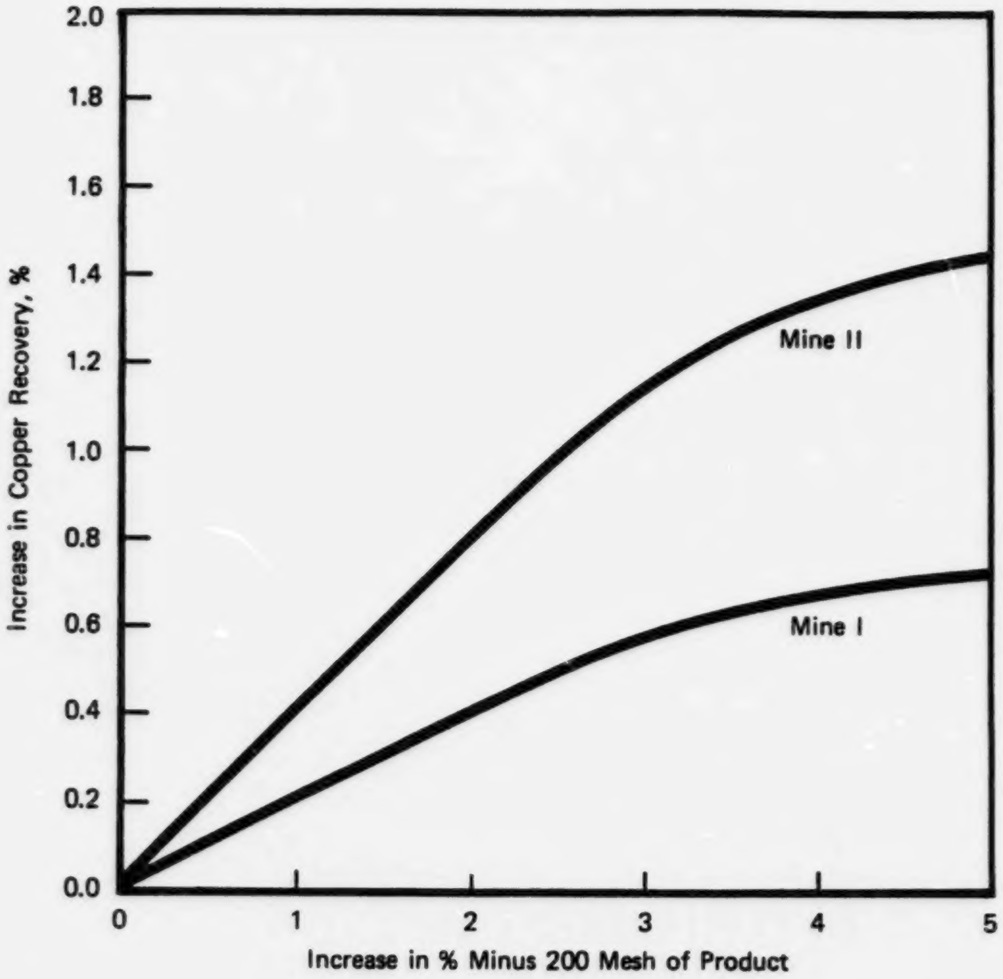


FIGURE XI. Illustration of extra copper recovery as a function of increased fineness of grind.

additive might still be justifiable by increased throughput, but the calculation of economic return is more complex because of cost allocation for extra mining, flotation chemicals, tailings disposal, etc. If the plant is operating near full capacity and extra throughput is desired, the chemical approach is generally quite easily justified. At less than capacity the increased throughput evaluation is more complex but at least should be compared to savings in power and steel required on a unit ton basis.

A method to predict the effect of chemical usage by scale-up from laboratory tests to industrial scale has been developed (11). It consists of estimating the approximate S curve for the industrial circuit in question. This was done using the analytical solution for the continuous plug flow case based on the following approximation (14),

$$1-P(x,\tau) = [1-P(x,0)]\exp(-S_i\tau).$$

The feed and output size distributions $P(x,0)$ and $P(x,\tau)$ were used to obtain the set of formal values of S_i using this equation. Then the proportional influence of grinding additive measured in a laboratory test was used to scale these S_i values, enabling an estimate to be made of the increase in feed rate (reduction in τ) at a constant product specification or of the increase in fineness at the same τ. Table 3 gives a typical set of full scale open circuit mill predictions for various slurry densities and chemical addition rates in the grinding of taconite ore.

Table 3 is essentially providing a "break-even" chart of potential production at similar product particle sizes. Therefore, for a given slurry density and chemical addition rate, if the feed rate is less than

Table 3. Prediction of rod mill performance on taconite in terms of increase in throughput with grinding additive, XFS-4272.

		Constant Particle Size Within Each % Solids Category			Constant Particle Size for all % Solids Categories		
Slurry Density wt % Solids	Chemical Addition Rate lbs/ton	Product Size %<35 U.S. Mesh	Feed Rate tph	Feed Rate % Increase	Product Size %<35 U.S. Mesh	Feed Rate tph	Feed Rate % Increase
84	0	40.0	410	-	40.0	410	Base
	0.4	40.0	410	0	40.0	410	0
	0.6	40.0	410	0	40.0	410	0
	0.85	40.0	410	0	40.0	410	0
	1.7	40.0	410	0	40.0	410	0
86	0	37.2	410	0	40.0	404	-1.5
	0.4	37.2	428	4.4	40.0	422	2.9
	0.6	37.2	437	6.6	40.0	431	5.1
	0.85	37.2	448	9.3	40.0	442	7.8
	1.7	37.2	478	16.6	40.0	471	14.9
88	0	34.8	410	0	40.0	392	-4.4
	0.4	34.8	450	9.8	40.0	430	4.9
	0.6	34.8	470	14.6	40.0	449	9.5
	0.85	34.8	494	20.5	40.0	472	15.1
	1.7	34.8	526	28.3	40.0	503	22.7

the figure stated, the product will be finer than the base case. If the feed rate is higher, the product quality will be coarser. Table 4 gives a set of open circuit rod mill results run on the same taconite ore to check the applicability of this simple approach for predicting changes in circuit output. It is seen that the predictions of Table 4 are qualitatively correct. Note that the slurry density used in rod milling tends to be 4-8 percentage points higher than in ball milling the same ore.

Inspection of Table 3 again, shows the type of trends discussed earlier on lab data. If the % solids is too low (or the mill is not in a viscosity limited state), adding a fluidity control chemical accomplishes little. In a range of higher % solids (usually Δ = 2-4%), the chemical is quite effective. If the % solids is too high (with the ore of Table 3, this is at 89%), the value of the chemical diminishes rapidly.

At Whitehorse Copper, exactly the same thinking is applicable except that in the normal Whitehorse operation (hence normal % solids region), adding the chemical gives a finer grind without any changes (e.g. 3% finer passing 200 mesh). Then as the feed rate is increased with the same chemical addition rate, about a 10% increase can be added while still giving the same product size as normal operation without chemical. Above this 10% at the same chemical level, the product gets coarser. For example, a 17% increase in throughput gave a 2.2% loss in product fineness at 200 mesh. Obviously, this decrease in fineness is much less than would be experienced at 17% without aid.

Conclusions

The use of grinding additives for the wet grinding of ores, based on efficient slurry modification under a high shear rate, has been shown to be

Table 4. Results of industrial rod mill with taconite run with additions of grinding aid.

				Rod Mill	
Time h		Rate tph	Polymer Addition XFS-4272 lbs/ton	Slurry Density wt % Solids	Product Size % <35 U.S. Mesh
Day 1					
0	Normal Operation	405	0	84	38.4
2	Aid Added	405	0.6	85	38.3
3.5	Higher Feed	425	0.6	86	42.7
4.7	Higher Feed	465	0.6	87	37.8
6.8	Higher Feed	485	0.6	89	34.2
7.1	Aid Off	485	0	-	*
7.5	Feed Off	000	0	87	22.5
8.7	Normal Operation	440	0	-	40.6
Day 2					
0	Normal Operation	440	0	87	41.1
1.7	Aid Added	440	0.4	87	40.2
2.5	Higher Feed	490	0.4	88	-
3.7	Steady-State	490	0.4	89	39.2
4.5	Aid Off	440	0	-	-
6.5	Normal Operation	440	0	-	45.3

* Pebble-sized particles appeared in mill product.

a predictable and logical phenomenon. The chemicals involved, e.g. XFS-4272, are generally low molecular weight anionic water soluble polymers having an appropriate hydrophobic balance so as to promote dispersion of dense slurries under a wide variety of pH, shear forces, water quality, and ore surface conditions. The data clearly suggests that more dense slurries interacting with tumbling media are slightly more efficient for actual particle breakage than the ordinary slurry densities when the mass transport and slurry viscosity limitations normally experienced with dense slurries are removed by the use of an additive. This is not to say that grinding additives based on other mechanisms are impossible or undesirable (e.g. strength modifiers). It appears that the chemical concept using XFS-4272 may be acceptable in some operations with on-site additive costs of 12-30 U.S. cents (1978) per ton of feed ore, giving a 6-10% increase in throughput at constant grind or a 2-4% finer grind of minus 200 mesh at constant throughput. The grinding results at Whitehorse Copper Mines clearly are positive and follow the engineering theory previously outlined rather closely. Also, no detrimental effects on flotation, waste disposal, etc. have been observed at Whitehorse Copper to date. In fact, in some mines the use of XFS-4272, when used in low dosages, has given higher copper recoveries. Further testing at Whitehorse Copper will be required to fully identify the economics involved.

Acknowledgments

The authors would like to express their appreciation to Basil Fee of Dow Chemical of Canada and Andy Mular of the University of British Columbia for the role they played in setting up the Whitehorse Copper Mine trial and subsequent encouragement and support of the grinding aid concept. Also special thanks should be given to Mel Katzer, John Sewell and Bob Hansen of Dow Chemical U.S.A. for their efforts in generating supportive laboratory data for grinding aid evaluation.

References

1. Snow, R. H., Chemical Engineer's Handbook, 5th Ed., H. R. Perry and C. H. Chilton, Editors, Published by McGraw Hill Book Co., 1973, Chapter 8, 12.

2. Rose, H. E., and Sullivan, R. M., "The Role of Additives in Grinding," Ball, Tube, and Rod Mills, Chemical Publishing Co., New York, 236-251, 1958.

3. Rehbinder, P. A., and Kalinkovskaya, N. A., J. Tech. Phys. (USSR), 2, 726-755, 1932.

4. Westwood, A. R., "Environment-Sensitive Mechanical Behavior," Metallurgical Soc. Conf., 35, Gordon and Breach Sci. Publishers, New York, 1-65, 1966.

5. Westwood, A. R., "Control and Application of Environment-Sensitive Fracture Processes," J. of Materials Sci., 9, 1871-1895, 1974.

6. Locher, F. W., and von Seebach, H. M., Ind. Eng. Chem. Proc. Des. Dev., 11, No. 2, 190-197, 1972.

7. Hartley, J. N., Prisbey, K. A., and Wick, O. J., "Chemical Enhancement of Ore Grinding Efficiency," Annual Intermountain Minerals Conference of the AIME, 1976, Vail, Colorado.

8. Klimpel, R. R., and Manfroy, W., "Computer Analysis of Viscosity Effects on Selection for Breakage and Breakage Distribution Parameters in the Wet Grinding of Ores," 14th Int. Sym. on Appl. of Computers in the Mineral Ind. (1966), S. Ramani, Editor, Published by SME/AIME, 1977, Chapter 16, 197-206.

9. Klimpel, R. R., and Manfroy, W., "Grinding Aids for Increased Throughput," Ottawa, Canada, January 1977. Proc. of 9th Annual Meeting of Canadian Mineral Processors, R. Bruce, Editor, Published by Can. Min. Proc., 1977, 206-218.

10. Klimpel, R. R., and Manfroy, W., "Development of Chemical Grinding Aids and Their Effect on Selections-for-Breakage and Breakage Distribution Parameters in the Wet Grinding of Ores," Proc. 12th Int. Min. Proc. Congress, Aug-Sept. 1977, Sao Paulo, Brazil.

11. Klimpel, R. R., and Manfroy, W., Ing. Eng. Chem. Proc. Design and Dev., 17, 518-523, Oct. 1978.

12. Austin, L., Powder Tech., 5, 261-266, 1971.

13. Austin, L. G., and Klimpel, R. R., Ind. and Eng. Chem., 56, No. 11, 18-29, 1964.

14. Herbst, J. H., Fuerstenau, D. W., Trans. SME/AIME, 241, 538-543 (1968).

Discussion of Paper by R. Kimpel

R. Bruce: How long will the effect last, such as when there is recycling of a stream ?

Reply: In copper circuits there are quite a few recycling flows and the benefits are most noticeable. One of the problems is the type of dispersant that is used. We advocate the polymeric type because very small amounts are needed. There is a variety of coat-dispersants that work well. With the inorganic type you use 6 to 10 times as much, which could amount to 4 or 5 pounds per ton. Reagent X is biodegradable. In flotation, such an inorganic dispersant as sodium silicate can ruin the circuit.

D. Doyle: Is Reagent X useful in dry grinding ?

Reply: We have at least four chemicals that can be used in dry grinding. It is a slightly different mechanism and engineering concept but it is well known.

K. Konigsmann: What is the cost of this reagent ?

Reply: It is an experimental product. In tests in the United States it cost 12 to 30 cents per ton to achieve a 10% increase in thruput or a 3% increase in grind.

Paper No. 18

LEACHING PRACTICE AT DENISON MINES
AN HISTORICAL REVIEW

E. LaRocque, Chief Metallurgist January 1979

INTRODUCTION

Denison Mines Limited has been processing uranium ore in the Elliot Lake area since May of 1957. The ore body is located in the Huronion sediments on the north limb of the Quirke Lake syncline. Minerals of metallurgical significance are brannerite, uraninite and monazite. Gangue minerals include quartz, chlorite, sericite, pyrite, pyrrhotite and rutile. Silica content of the ore is greater than 75 % and pyrite-pyrrhotite content around 7 %. Specific gravity of the ore is 2.75.

The Denison flowsheet is based on sulphuric acid leaching to extract the uranium values from the ore followed by liquid-solid separation to wash the uranium-bearing solution from the barren ore. The resultant pregnant solution is concentrated and purified by ion exchange followed by two-stage precipitation with ammonia to produce an ammonium yellow-cake.

Elliot Lake ores require high acid concentrations and high temperatures to extract the uranium from brannerite and monazite. Uraninite will leach at low acid concentrations as long as oxidizing conditions exist. The whole ore must be treated so that these conditions do exist.

This paper provides a description of the leaching process equipment as originally installed and major changes in that equipment to date. Process control refinements to economically improve extraction are also discussed.

ORIGINAL INSTALLATION

The original leaching plant was designed to incorporate a recycle acid leach process. The ore was ground at 65 - 70 %, -200 mesh, dewatered in neutral thickeners and further dewatered on neutral disc filters. The filter cake was repulped with hot recycle acid and pumped to the leaching plant.

The leaching plant was comprised of three circuits, each one containing two 15' x 35' air agitated pachucas and six 32' x 30' centre column airlift wood stave agitators.

Leach discharge pulp was filtered on acid disc filters. The acid solution was passed through a Karbate heat exchanger and used to repulp the neutral disc filter cake. The slurry was then pumped to the leaching plant. The acid filter cake was repulped, partially neutralized to pH 1.8 and processed through the liquid-solid separation section.

Moisture on both the acid and neutral disc filters was around 15 %. When the filters were balanced properly, little or no acid needed to be neutralized and all the acid could be used to repulp the neutral disc cake. Maintenance costs were high on the disc filters and the heat exchanger. Sector bags were replaced at the rate of 2,500 per month on the acid filters and 1,500 per month on the neutral filters. Studies showed that the savings in sulphuric acid, lime and steam more than offset the maintenance and operating costs.

The airlift agitators continually sanded up. Grinding was controlled so there was less than 1.5 % + 48 mesh material in the ore for this reason. Animal glue was added to help suspend the sands, at times as much as 2 pounds per ton. In 1958, a polymer was used instead of the animal glue. The density of the leach was kept as high as possible after 1958. As the density increased, sanding problems lessened and polymer addition ceased. Maintenance and down time were high for repairs to the 316 stainless steel columns, rakes and chains.

PROCESS CONTROL

(1) Sulphuric Acid Addition

Sulphuric acid (93%) was added to the first pachuca in each circuit. Addition was manually controlled. The operator titrated for free acid concentration and opened or closed a control valve as needed. During 1957 and 1958, the control level was 75 grams per litre free sulphuric acid. Acid levels were lowered to around 60 grams per litre in 1959. Acid control was erratic, particularly if acid consumers were present in the ore or if the operator was having other operating problems.

(2) Sodium Chlorate Addition

Sodium chlorate was added as a 10 - 15 % solution to the discharge of the second pachuca. There had been a great deal of discussion in the Elliot Lake camp as to the levels of ferric iron that should be maintained for optimum extraction. Total iron in solution was between 7 and 8 grams per litre.

Uranium in the quadravalent state U(1V), as UO_2, is insoluble in H_2SO_4. It is necessary to have ferric iron plus an oxidant to convert the quadravalent U(1V) to the hexavalent U(V1). The uranium can then be leached with sulphuric acid.

From 1957 to 1960, sodium chlorate addition was controlled so as to result in a level of 4 grams per litre ferric iron in the first agitator. Chlorate consumption averaged 3.5 pounds per ton of ore for that period.

In 1961 chlorate was added so as to result in 1 gram per litre ferric iron. Consumption of sodium chlorate dropped to 1 pound per ton. As shown in Table 1, the leach tail did increase, but there was a considerable drop in the consumption of sulphuric acid, lime, and sodium chlorate. It should be noted that this was the same period of time that the U.S.A.E.C. had indicated they would not renew their option to purchase uranium concentrates and production economies were becoming ever more important.

TEMPERATURE CONTROL

The wood stave agitators were constructed of 6" B.C. fir and were not recommended for temperatures greater than 45°C. The plant was operated at this temperature until 1960. Testwork showed that extraction was better as temperature was increased. From March 1960 to October 1961, the temperature was increased in increments from 45° C to 72° C. Referring again to Table 1, it might appear that the increase in temperature in itself did not improve extraction but overall reagent savings are attributable to the higher temperatures.

The elevated temperatures did, however, leach the lignin from the wood. Beginning in May, 1963, agitators were taken out of service and dismantled and by 1965, it became apparent that a new leaching plant would have to be installed. A decision was made to discontinue the recycle acid leach process and change to a straight through leach in 22' x 50' air agitated pachucas. A new addition was made to the mill and 12 pachucas were installed. Operation of the pachuca plant was initiated in August, 1966.

DENSITY CONTROL

The leaching plant had been designed to operate at 70% solids in order to make full use of the recycle acid. Densities were lower than this until 1959 and caused sanding problems in the agitators as well as high reagent consumption. When the plant operated at 70 % solids, there was no problem with sanding and reagent consumption improved.

PROCESS CONTROL IMPROVEMENTS 1957 -1966

Grinding

In January 1960 the conversion in the grinding plant from ball milling to pebble milling was complete. It had been expected that the exclusion of the 2.5 pounds per ton of steel from the balls would reduce the consumption of sulphuric acid and sodium chlorate. There was no apparent change in reagent consumption and leach tails in fact were higher in 1960 than they had been in 1958 and 1959, up from 0.14 pounds U_3O_8 per ton to 0.16 pounds . There was of course a saving in the cost of the steel balls. The final grind before and after the pebble milling change remained about the same.

+48 M	+65 M	+100 M	+150 M	+200 M	-200 M
0.2	3.4	12.3	11.0	11.2	61.9

This grind was maintained until 1966 when the pachuca plant was put on stream.

Automatic Controls

During the latter part of 1962 automatic controls were installed to control sulphuric acid, sodium chlorate and steam to the leaching plant.

(1) Sulphuric Acid

Conductivity probes were installed in the lead pachuca and controlled the acid levels within 2 grams per litre. This smoothed out the operation and insured the same acid level when acid consumers were processed through the plant. While there was an increase in acid consumption the leach tail was reduced from .17 pounds U3O8 per ton in 1961 and 1962 to 0.15 pounds per ton in 1963. A computer study completed in 1963 had shown that better extraction could be attained by adding all the acid and heat as close to the head end of the leaching circuit as possible. The acid control was moved to the neutral disc repulper in June, 1963 and extraction did improve.

(2) Sodium Chlorate

Redox controllers were installed to control the redox at 420 millivolts. The addition smoothed out and there was a reduction in sodium chlorate consumption from 1 pound per ton to about 0.6 pounds per ton.

Temperature Control

Automatic steam control valves were installed and smoothed out the temperature control.

MISCELLANEOUS

The original columns in the agitators were type 316 stainless steel. The higher temperatures combined with the chloride content of the solution caused severe pitting corrosion on the columns. Beginning in 1960 the columns were replaced with rubber lined and covered pipe.

While it might appear from Table 1 that extraction improved as the tonnage rate decreased in the mid 60's, this is not necessarily so. When the recycle acid leach plant was closed out in August of

1966 retention time in the plant had also been decreased because of the loss of agitators. During the last few months of operation of this plant there were three circuits, each one comprised of two pachucas and one agitator, down from the original circuit which was comprised of three circuits each one containing two pachucas and six agitators.

CURRENT LEACHING PROCESS

The recycle acid leach circuit was phased out in 1966, and a new pachuca plant constructed on the north side of the metallurgical plant.

The leaching circuit consisted of two parallel lines of six pachucas each. The pachucas were 22.5 feet in diameter and 50 feet high with a 60° bottom cone. The pachucas were rubberlined to withstand the high temperature, high acid concentrations and oxidation-reduction potential conditions associated with the leaching of Elliot Lake ores. Air was injected at 40 - 44 psig through three, one-inch 316 stainless steel valves located at 27" centres in the conical bottom section.

Six of the acid disc filters from the old filter section were installed above the pachucas and used for dewatering neutral thickener underflow prior to discharge into the first pachuca in each line.

The reasons for installing pachucas rather than duplicating the recycle acid leach circuit were:

(1) Higher densities could be maintained in pachucas thus reducing sulphuric acid consumption.

(2) Higher temperatures could also be maintained without permanent damage such as had occurred in the wooden agitators.

(3) It would not be necessary to maintain such a fine grind. Grinding costs could be reduced and settling and filtering improve.

(4) Maintenance costs would be reduced.

(5) It was felt that the air injection would provide enough oxidation and sodium chlorate addition would not be necessary.

It became apparent immediately that the air pressure (40 - 44 psig) was not sufficient to agitate the pachucas at the high densities (77 % solids) that would allow a reduction in sulphuric acid consumption.

When the densities were high, the air flow dropped, redox potential dropped, uranium extraction suffered and sulphuric acid consumption was low.

As densities were lowered towards 70 % solids, the air flow increased, redox potential went up, extraction improved, but sulphuric acid consumption increased.

The pachucas had to be operated at densities between 70 and 72 % solids to maintain an acceptable extraction. There was a seesaw battle for years trying to find a balance between extraction and acid consumption (Table 11).

From August, 1967, through 1968, the sulphuric acid addition was controlled at 44 grams per litre in the lead pachuca. This was raised to 48 and then to 52 grams per litre in 1968 to improve extraction. The acid levels were raised to 60 grams per litre in 1969 and maintained at that level until 1973.

As much air as possible was added to the first two pachucas in line. The best air flow possible at 72 % solids was 380 c.f.m.

Temperature was controlled during this period at 70° C.

In November of 1973, even though the leaching controls were the same as they had been for years, the uranium in the leach tail suddenly rose from an average of 0.16 pounds per ton to over 0.40 pounds per ton. When the temperature was raised to 75° C, and the acid levels to 75 grams per litre, the leach tail came back down to 0.16 pounds per ton.

Test work had shown that the addition of "Royal Barren". the first portion of the ion exchange barren, would improve extraction. This solution contains 6 to 10 grams per litre nitrate and it had been known for years that nitrate would aid uranium extraction. This solution which has been used to repulp the disc filter cake since May of 1974, reduced the solid tail to less than 0.14 pounds U3O8 per ton for the balance of 1974.

Booster air compressors were installed in 1974 to raise the pressure to 52 psig. Air flows were raised to 450 c.f.m. in the lead pachucas and densities have been maintained at 75 % solids since that time.

In 1974 the capacity of the metallurgical plant was raised to 7,100 tons per day. As part of that expansion ten more pachucas were added. The present leaching plant now consists of two circuits, each containing eleven pachucas.

From the experience of the high tails in late 1973 samples were collected from underground developement headings, and a comprehensive program of leaching tests carried out. Results of these tests showed that a significant percentage of future planned production required sodium chlorate addition to maintain or improve extraction. At additions of up to 300 pounds per ton of sulphuric acid and a temperature of 80°C, extraction was only 50 %. At current (1975) acid and temperature levels, the addition of 1 to 2 pounds of sodium chlorate resulted in extractions exceeding 95 %.

The sodium chlorate system was overhauled and full scale plant trials of several months duration were carried out during 1976 and 1977. Results were encouraging and sodium chlorate was used throughout 1978. Leach tails averaged 0.11 pounds per ton. Chlorate consumption was not recorded in Table 11 for 1976 and 1977 as it was only being added on a test basis for that period.

The current leaching control parameters are:

(1) Sulphuric acid levels of 75 grams per litre in the first pachuca.

(2) Temperature - 80°C in the lead pachuca. No heat in the last five pachucas.

(3) Sodium chlorate addition controlled to give a redox of 450 millivolts in the second pachuca.

(4) Density - 75 % solids.

(5) Air flow of 450 c.f.m. in the lead pachuca, 350 - 400 c.f.m. in balance of circuit.

These conditions are resulting in a leach tail of between .08 and .1 pounds U3O8 per ton. Test work on development headings will continue to try and foresee any necessary changes for future production.

RECYCLE ACID LEACH CIRCUIT

TABLE 1

Year	Tonnage	Head Assy Lbs. U_3O_8/Ton	% Loss Leach Tail	% Extraction	Leach Tail Lbs. U_3O_8/Ton	Reagent Consumption lbs./Ton		
						H_2SO_4	Lime	Sodium Chlorate
1957	2,676	2.63	7.95	92.05	.209	79.5	44.7	3.4
1958	5,101	2.46	5.37	94.63	.132	91.3	48.3	4.0
1959	5,606	2.56	5.53	94.45	.142	71.2	42.6	2.8
1960	5,787	2.70	6.08	93.93	.164	70.0	42.0	3.3
1961	5,827	2.85	6.08	93.93	.173	62.4	32.8	1.1
1962	5,680	2.88	6.00	93.99	.173	62.7	32.4	0.9
1963	4,444	3.34	4.66	95.33	.156	69.6	38.3	0.6
1964	3,573	3.14	3.94	96.05	.125	67.4	37.5	0.7
1965	2,624	2.93	4.20	95.80	.123	66.0	37.2	0.6
1966	2,922	2.86	4.55	95.45	.130	71.3	41.0	0.4

CURRENT LEACHING CIRCUIT

TABLE 11

Year	Tonnage	Head Assay lbs.U3O8/Ton	% Loss Leach Tail	% Extraction	Leach Tail lbs.U3O8/ton	Reagent Consumption lbs./ton		
						H_2SO_4	Lime	Sodium Chlorate
1967	3,416	3.07	4.95	95.05	0.152	87.0	49.5	-
1968	3,858	3.07	5.42	94.58	0.166	80.3	47.0	-
1969	3,515	3.43	4.79	95.21	0.164	76.1	44.7	-
1970	3,497	3.15	4.83	95.17	0.152	69.6	44.0	-
1971	4,128	3.20	4.88	95.12	0.156	72.3	48.3	-
1972	4,300	2.87	5.61	94.39	0.160	69.8	47.3	-
1973	4,287	2.57	6.36	93.64	0.164	63.2	46.9	-
1974	3,983	2.33	6.30	93.70	0.147	63.0	44.9	-
1975	3,818	2.30	5.72	94.28	0.132	67.8	52.9	-
1976	4,741	2.16	4.95	95.05	0.107	79.7	58.5	-
1977	5,884	2.07	5.14	94.86	0.107	80.4	56.9	-
1978	7,092	2.15	5.07	94.93	0.109	80.7	53.7	1.7

Discussion of Paper by E. LaRoque

E. Marcotte: What is the phenomena in the use of Royal Barren with the neutral filter cake before leaching ?

Reply: The idea is to utilize the unused nitrates from the ion exchange circuit. The nitrate level in this solution is about 6 to 10 gpl. For many years we neutralized the Royal Barren and then added the neutralized cake to the leaching circuit but it created a lot of problems in settling and filtering.

H. McCreedy: Where are the conductivity probes located ? Do you notice any corrosion or errosion in the high acid circuit ?

Reply: We installed stainless steel steam injectors in the first pachucas in May, 1973 and they lasted only one week because of the high reducing condition that existed there. The EMF in the first pachuca runs from minus 100 mv to plus 500 mv indicating a great variation in oxidation-reduction conditions from time to time. The conductivity probe is installed between the first and second pachucas. Sodium chlorate is added to the second pachuca.

Paper No. 19

RIO ALGOM LIMITED

Mining Division

OPERATION OF ENVIRO CLEAR THICKENERS IN SERIES AS CCD UNITS

By Lloyd W. Hamel
Rio Algom Limited

For Presentation at the Canadian
Mineral Processors Meeting
January 25, 1979

SUMMARY

Two 35 foot diameter enviro clear thickeners were tested in series as CCD units from June until October, 1978. Up to shutdown in July, we ran with automatic bed level control via the level probes and underflow pinch valves. Despite extensive efforts, we were unable to eliminate the surges created between the units by the opening and closing of the pinch valves. Early in July, the best rake speed and gap setting were determined. After shutdown, the enviro clears were run by manually controlling the pinch valves. These efforts were successful.

A few mechanical problems were encountered while operating the thickeners. New units should incorporate more support for the drive coupling and a rubber feed seal. A few modifications of the instrumentation control scheme will be required before the units can be fully automated. A less sophisticated automatic flocculant addition system will accomplish the desired results. Control of the underflow withdrawal via the pinch valve should be investigated further. Our best results were obtained by keeping the feed volume to the No. 1 enviro clear constant. Surges were controlled by opening or closing both underflow valves proportionately.

Capacity of the enviro clears is dependent of many variables. The ore treated during the test program was much finer than that reported on in R & D Project Report 76-02. Conditions were never steady enough to determine best flocculant dosage or addition point(s). The lowest soluble losses were obtained at very high wash rates. The underflow densities drop but the soluble losses also drop. Regression equations were developed in order to assess the effect of the various parameters. Within the range studied, these equations will predict the performance of enviro clear thickeners as CCD units on Elliot Lake ores. Our results show that the enviro clear thickeners are a viable alternative to the conventional CCD thickener.

1. INTRODUCTION

Since 1975, Rio algom has become progressively committed to the use of enviro clear thickeners. Results from treatment of mine-water, first and sixth stages of CCD and neutral thickening pulps on laboratory and pilot plant scale were encouraging enough to decide to purchase a full scale, 30 foot diameter enviro clear thickener for mine water treatment. Prior to installation for mine water clarifying, the enviro clear was tested as a sixth stage CCD unit. Only a little over a month was available to run the unit. The next step was to run three units in series treating CCD pulps. It was hoped to finish this test program before final commitment had to be made on the Bank C expansion of the CCD circuit. Construction and instrumentation delays in completing the installation of the enviro clears squashed any hopes of using enviro clears for Bank C.

The enviro clear in-series test program commenced in late May of 1978 with two units. High solution losses on the new conventional Bank C thickeners had necessitated repiping the third enviro clear as the sixth stage of Bank C. Testing continued until the end of September/78.

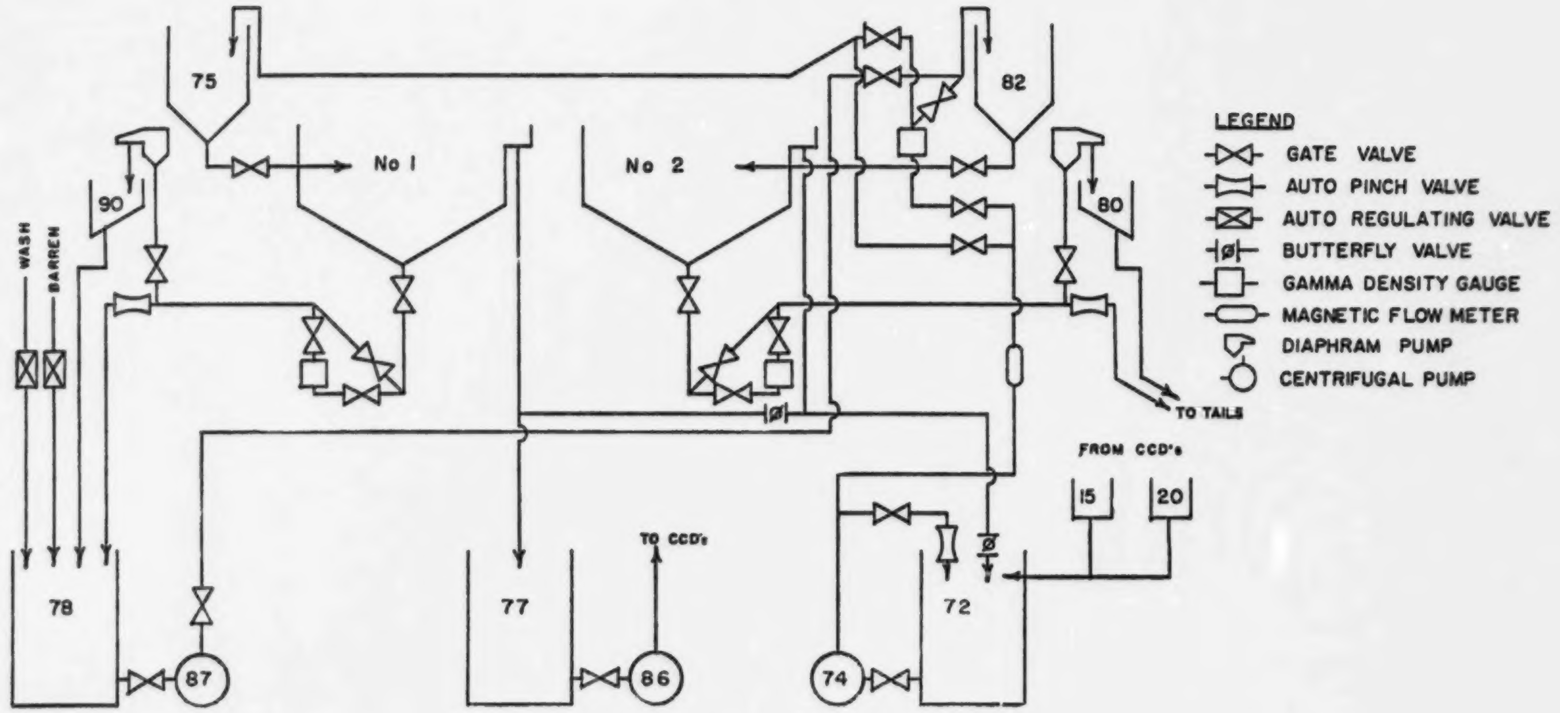

75
82
No 1
No 2
90
80
WASH
BARREN
LEGEND
GATE VALVE
AUTO PINCH VALVE
AUTO REGULATING VALVE
BUTTERFLY VALVE
GAMMA DENSITY GAUGE
MAGNETIC FLOW METER
DIAPHRAM PUMP
CENTRIFUGAL PUMP
TO TAILS
FROM CCD's
15
20
TO CCD's
78
77
72
87
86
74

FIGURE I PROCESS SCHEMATIC

2. DETAILS

2.1 Process Piping

A schematic of the two stage series enviro clear test is shown in Figure I. Underflow from A5 and/or B5 conventional CCD thickener flows by gravity to box 72. Overflow from No. 2 enviro clear combines with this pulp to become No. 1 enviro-clear feed. Pump box 72 has a bubble tube level control system to ensure no downstream surging occurs. A magnetic flow meter and a gamma density gauge are mounted on the line for determination of the amount of dry tons of solids per hour being fed to the circuits. Feed to No. 1 enviro clear first enters a de-aeration tank and then gravitates to the No. 1 enviro clear. Overflow drains into box 77 and then is pumped back to the fifth stage feed pump boxes of Banks A and B. Underflow is either withdrawn via a 2 speed diaphragm pump up to box 90 and then drains to box 78 or drains directly to box 78 through a pinch valve. Wash and Barren are combined with the underflow to become feed to the No. 2 enviro clear. The feed is pumped up to a deration tank, before gravitating to the No. 2 Enviro feed well. Overflow drains into box 72 and is combined with underflow from the fifth stage of Banks A and B to become No. 1 enviro clear feed. Underflow is either pumped or drained to tails. Flocculant is supplied at .05% strength from a steady head system through a rotameter with manual adjustment or by a moyno metering pump. Flocculant is added to any one or more of five addition points along the feed pipe and into the feed well.

2.2 Instrumentation

One of the main reasons the enviro clear series test program took so long to get going was due to the instrumentation calibration. Quirke instrument technicians were hard pressed to keep up with the operating plant and current expansion, let alone the enviro clear series test. In March of 1978, a private contractor was hired to complete all the construction instrumentation. Special calibrators had to be ordered from Foxboro. It took until June of 1978 to calibrate all the instrumentation.

The instrument control scheme used for the test includes:

2.2.1 Dry Tons Per Hour

Figure 2 contains a schematic showing how the dry tons per hour of feed solids is determined in the field. Total Flow is measured by a 0-1200 IGPM Foxboro magnetic flow meter. From the flow meter, a 4-20 ma signal is sent to the control panel where it is changed to 0-10 volts. Density of the feed is measured by an Ohmart gamma gauge. The density transmitter also sends a 4-20 ma signal to the control panel which is changed to 0-10 volts at the panel. Range of the gamma gauge is between 1187 and 1395 S.G. A multiplying board then calculates the dry tons per hour of feed according to the equation

$$\text{DTPH} = \frac{(.992\ \text{DENSITY} + 8.92)\ \text{FLOW}}{10}$$

For example:

Flow = 600 IGPM - 12 ma - 5 volts
Density = 1300 S.G. - 12.7 ma - 5.43 volts
DTPH = 0-120 TPH 0-10 volts

$$\text{DTPH} = \frac{(.992 \times 5.43 + 8.92)\ 5}{10} = 7.15 \text{ volts}$$

$$= 85.8 \text{ TPH}$$

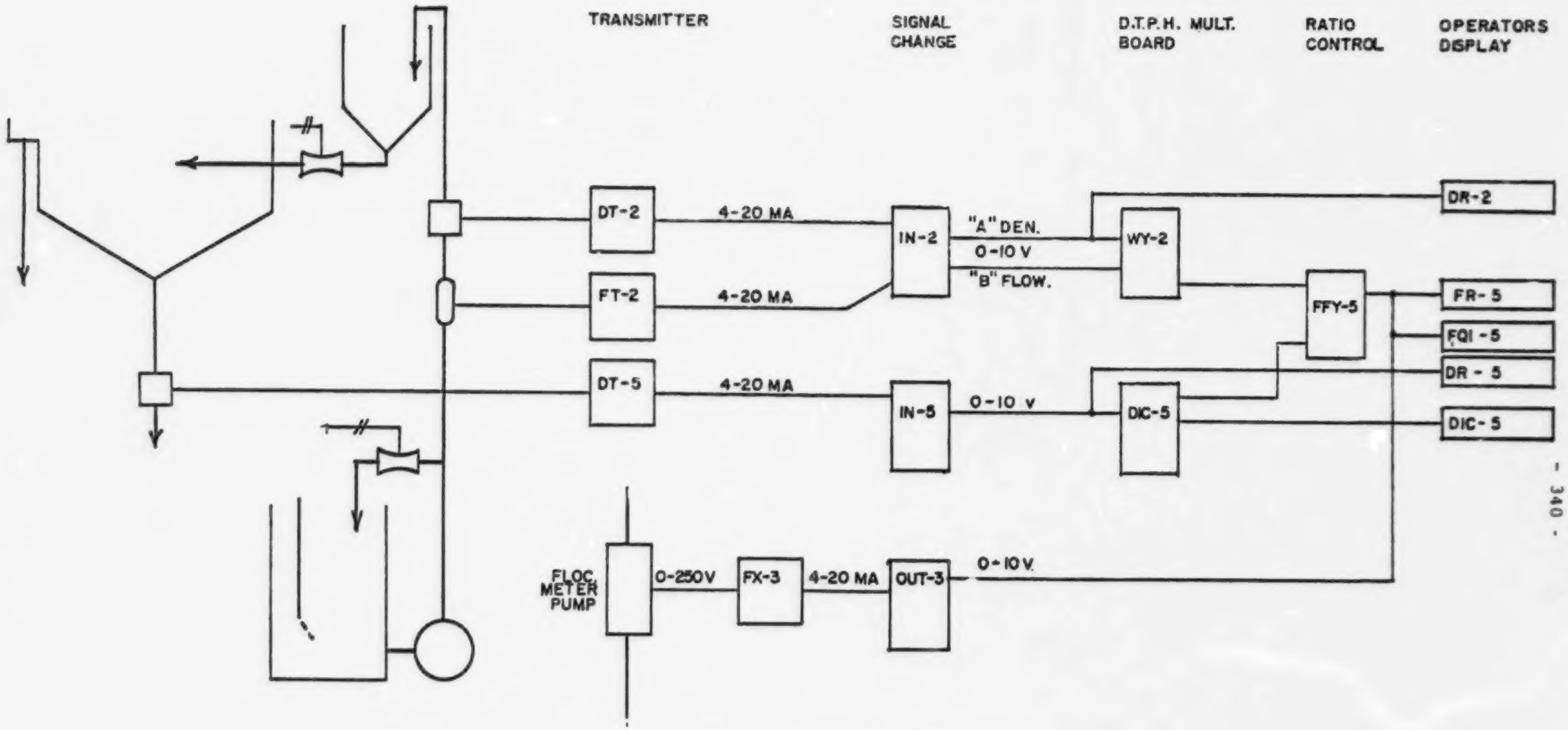

TRANSMITTER
SIGNAL CHANGE
D.T.P.H. MULT. BOARD
RATIO CONTROL
OPERATORS DISPLAY
DT-2
FT-2
DT-5
4-20 MA
4-20 MA
4-20 MA
IN-2
IN-5
"A" DEN.
0-10 V
"B" FLOW.
0-10 v
WY-2
DIC-5
FFY-5
DR-2
FR- 5
FQI - 5
DR - 5
DIC- 5
FLOC METER PUMP
0-250V
FX-3
4-20 MA
OUT-3
0-10V.

FIGURE 2 INSTRUMENTATION CONTROL

Actual dry ton per hour

Total weight/min of slurry

$$600 \text{ IGPM} \times 1.3 \frac{\text{gm}}{\text{cc}} \times \frac{10 \text{ lb.}}{\text{IG}} = 7800 \text{ lbs./min}$$

% solids of pulp 1300 SG

$$\% \text{ solids} = \frac{SG\ (SP-1)}{SP\ (SG-1)} \times 100\% = \frac{2.7(1.3 - 1)}{1.3(2.7 - 1)} \times 100\% = 36.65\%$$

Then total solids/min

= 7800 X .3665

= 2858.8 lbs/min

= 85.8 DTPH checks with instrument calculation.

The dry ton per hour of feed signal is used in determining the amount of flocculant to feed the enviro clears.

2.2.2. Underflow Density

Ohmart gamma gauges are used to monitor the underflow density from the enviro clears. The range of the units is set for 1400-1800 S.G. The 4-20 ma signal from the density transmitter is changed to 0-10 volts at the control panel. The underflow density is displayed on the operator's panel.

2.2.3. Automatic Flocculant Addition

Automatic flocculant addition is based on the dry tons per hour of feed and a modification from the underflow density signal. The density controller signal used in calculating the flocculant signal uses the density signal voltage plus or minus "n" volts.

For example:

U F Density : 1600 - 8 ma - 5 volts

U F Density set point: 1650 → change of density signal .5 V

DTPH : 85.8 - 7.15 volts

Then:

$$\text{Flocc Signal} = \frac{5.5 \text{ V} \times 7.15 \text{ V}}{10} = 3.93 \text{ V}$$

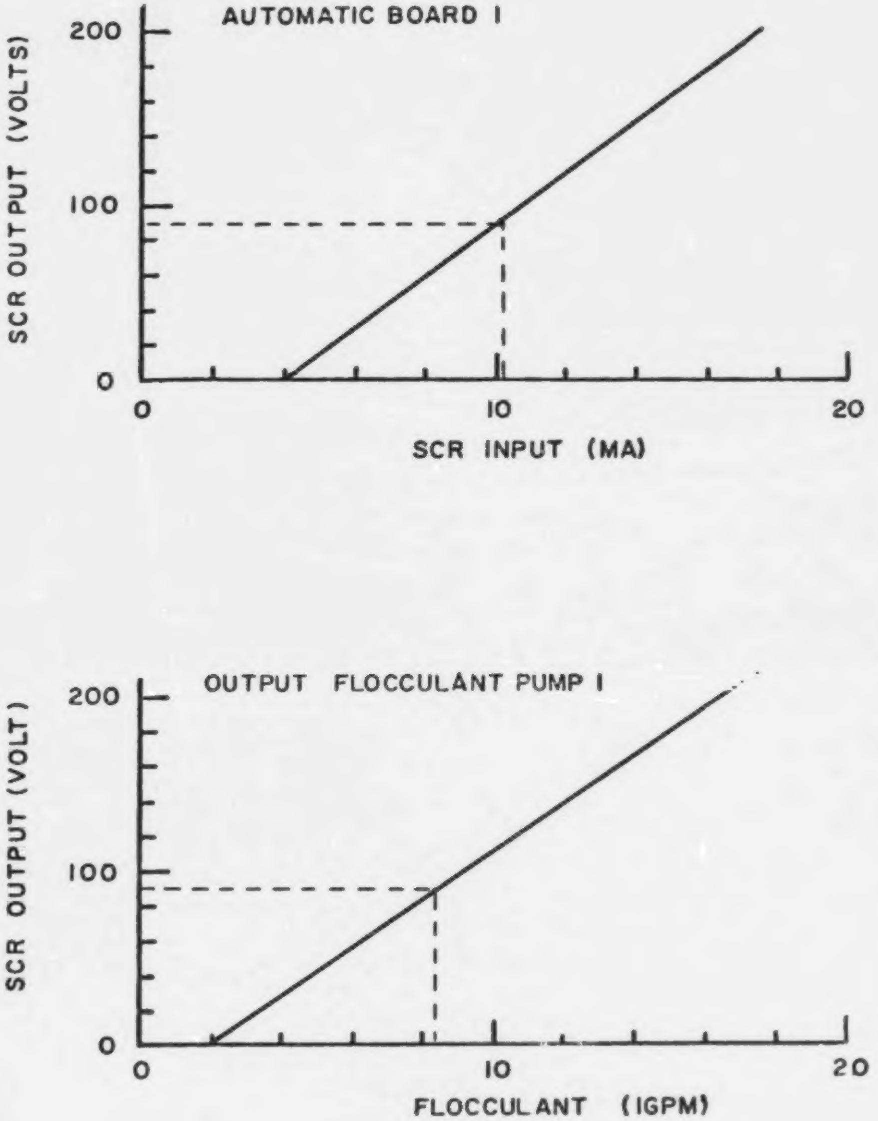


FIGURE 3 AUTOMATIC FLOCCULANT ADDITION

The flocculant signal is then changed to a 4-20 ma signal and is sent to the flocculant SCR controller. This controller is calibrated to change the 4-20 ma signal to 0-180 volts input to the flocculant metering pump. Flocculant delivered is between 0-15 IGPM. These relationships are shown graphically in Figure 3. In our specific example the flocculant dosage is:

$$\text{Flocc dosage} = \frac{\text{Flocc IGPM X STRENGTH X 6}}{\text{DTPH}}$$

Flocc signal → 3.93 V - 10.29 ma - 91 volts - 8.4 IGPM
DTPH 85.8
Flocc Strength 0.05%

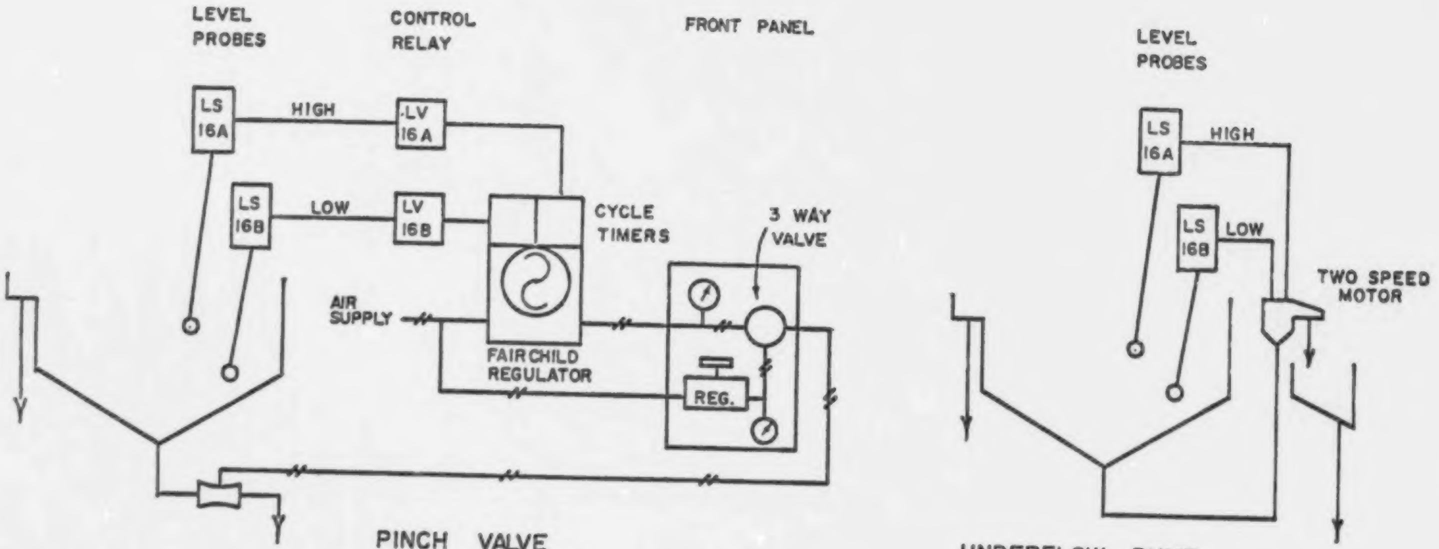
$$\text{Flocc dosage} = \frac{8.4 \text{ X } .05 \text{ X } 6}{85.8} = .029 \text{ lbs/ton}$$

Calibration of the automatic flocculant system took over four months to complete. Very little information was made available to determine how the system was designed to work.

2.2.4. Underflow Removal - Bed Level Control

Indication lights are used in conjunction with the ultrasonic probes to determine the relative bed level height. A yellow light corresponds to the very high level probe. A red light denotes a low level and a green light a high level. Operators are instructed to keep the bed level between the low and high level probes (red and green lights).

Two options are available for controlling the bed level. Both options are shown in Figure 4. The first system utilizes a two-speed diaphragm pump and the high and low level probes. Consider that the bed level has dropped past the low probe. The low pump speed will kick in and stay on until the bed level reaches the high probe. Then, the high pump speed kicks in and stays on until the bed level drops below the low probe. The pump speeds can also be manually chosen.


LEVEL PROBES
CONTROL RELAY
FRONT PANEL
LS 16A
HIGH
LV 16A
LS 16B
LOW
LV 16B
CYCLE TIMERS
3 WAY VALVE
AIR SUPPLY
FAIRCHILD REGULATOR
REG.
LEVEL PROBES
LS 16A
HIGH
LS 16B
LOW
TWO SPEED MOTOR

PINCH VALVE

UNDERFLOW PUMP

FIGURE 4 AUTOMATIC BED LEVEL CONTROL

The alternative option makes use of a Galigher Delta style pinch valve. Automatic bed level control is also by means of the high and low level probes. Control range on the Galigher valve is 4-12 psi. To provide some sort of proportional type control, ON/OFF cycle timers are incorporated along with Fairchild controllers.

Consider that the bed level has dropped past the low probe. The pinch valve's closing sequence then begins. The electrical circuit remains OFF for up to a minute and then ON for up to 10 seconds. When the electrical circuit is energized, the motor on the Fairchild controller screws down into the pressure regulator, resulting in a higher air pressure being delivered to the pinch valve. The OFF/ON sequence with the incremental increase in valve pressure continues until either the maximum pressure on the valve is obtained or the bed level rises past the high probe. Once the bed rises past the high probe, the valve opening sequence begins with incremental decreases in valve pressure. Valve pressure can also be manually chosen.

2.2.5. Rake Torque

Each one of the enviro clears came equipped with a Speedy Vickers HAS 7 hydraulic drive with a 10 h.p. electric motor. The arrangement of the instrumentation used to monitor and provide cut out protection from high torques is shown in Figure 5. The high torque alarm is set for a pump discharge pressure of 800 psi. High torque alarm is indicated in the field by an annuciator and on the operator's control panel by an annuciator and a flashing light. If the torque continues to rise, at 1500 psi, the pre-set cut out shuts off the electrical drive motor.

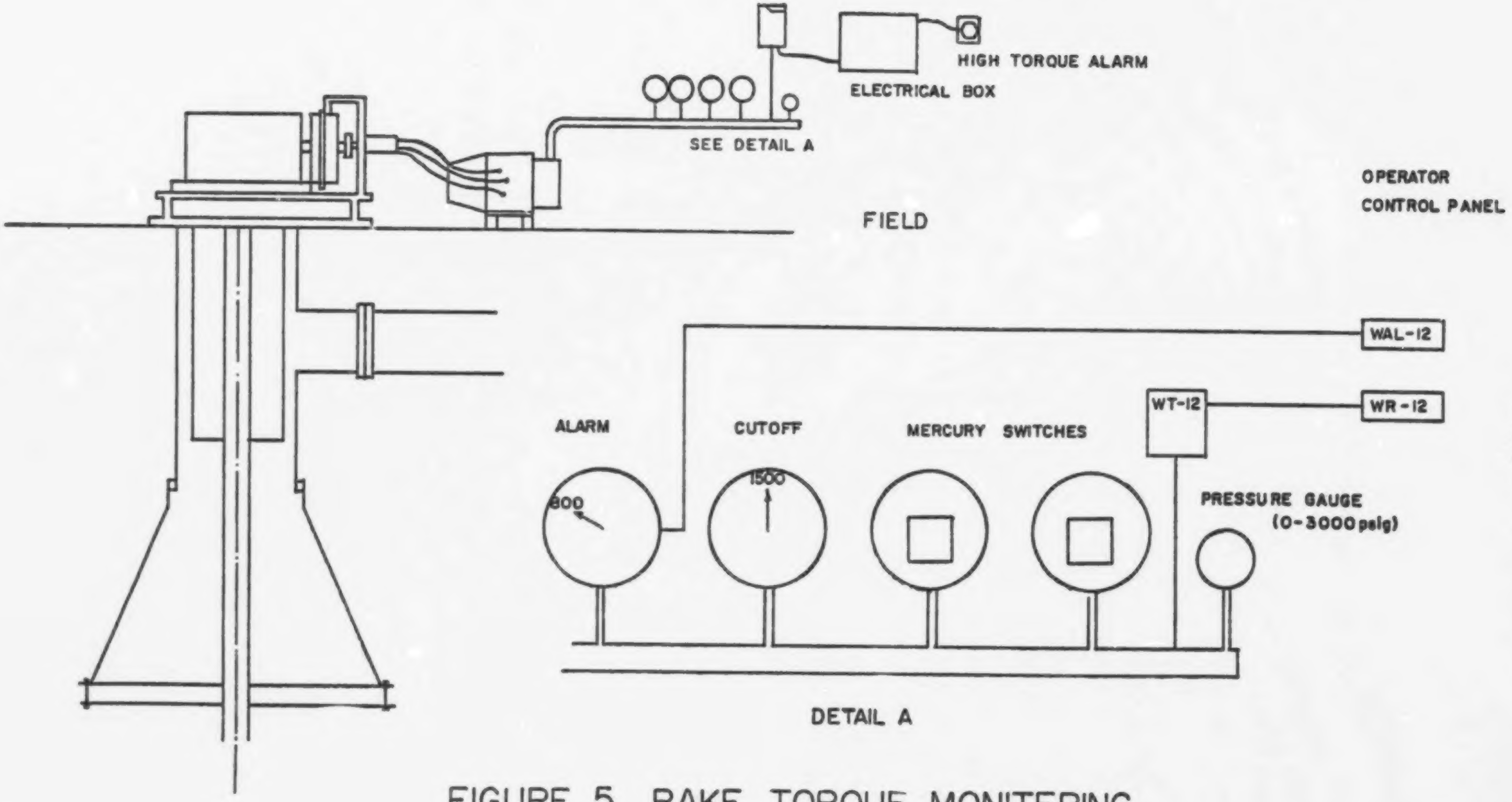

HIGH TORQUE ALARM
ELECTRICAL BOX
SEE DETAIL A
OPERATOR
CONTROL PANEL
FIELD
WAL-12
WR-12
WT-12
ALARM
CUTOFF
MERCURY SWITCHES
PRESSURE GAUGE
(0-3000 psig)
800
1500
DETAIL A

FIGURE 5 RAKE TORQUE MONITERING

3. TEST PROGRAM

The test program finally started in late May, when all of the instruments were finally calibrated. We were originally scheduled to run the program until the annual shutdown in late July. Without a six-month test program, we were forced to make a concentrated effort to determine the operating characteristics and instrumentation requirements in running enviro clear thickeners as CCD units in series. Process Development personnel were utilized to operate the units on a 24 hours per day, 7 days per week basis. Our original objectives were:

1. To run the enviro clear thickeners at the same conditions as the conventional thickeners.
2. Then run with increased wash.
3. Finally run with increased tonnage.

During the first week in late May, our efforts to accommodate swings in feed tonnage from CCD were fruitless. In one eight hour shift, the tonnage could vary from 30 to 80 T.P.H. Weirs were installed in A5 and B5 underflow pump boxes. We tried to hold the tonnage to the enviro clears constant and to gather data for a range of different wash rates. Some success was realized, but surges, created by the underflow pinch valves changing from the wide open to closed position, often overloaded the enviro clears. The difference in height between the high and low level probes was decreased from 2" to 1" to minimize the surge effect. Many combinations of ON and OFF cycle times were tried. Even though the final arrangement was better, surges still plagued the operation. We did not have the opportunity to try and perfect use of the two-speed diaphragm pump control system. The mill mechanical department had removed the gear box for the No. 1 unit for an emergency and never replaced it.

During a surge, as much as 700+ IGPM could be fed to an enviro clear. When the units were installed, the distribution gap was set at 1 5/8" for a nominal flow of 550 IGPM. This corresponds to a pulp exit velocity of 46 fpm. Technical representatives from Enviro Clear Corporation recommend setting the gap to obtain an exit velocity of 30-35 fpm.

3. TEST PROGRAM (Cont'd.)

We decided to open the gap to 2" which corresponds to an exit velocity of 47 fpm at 700 IGPM. In order to obtain further data, we ran various combinations of flow/gap settings.

Next, we investigated the effect of rake speed on the underflow density and torque. Enviro Clear's technical representative in New Jersey commented that they usually set the units to run at a peripheral rake speed of 1 fps. This corresponds to about 1/2 rpm in our case. He also added that you will get higher underflow densities and torques by operating at slower speeds.

The ore treated during the month of July was extremely fine. Problems in the conventional CCD circuits necessitated cutbacks of enviro clear wash on numerous occasions. During July, we continued to gather data at different tonnage and wash rates. Several attempts were made to treat 70+ DTPH. But on each occasion, the enviro clears overloaded and we were forced to cut back.

An Enviro Clear Corporation representative, Mr. W. Sackett was brought up to view our operation just before shutdown. During our discussions about operation of the enviro clears at other properties, namely Anaconda's, Grant's, New Mexico and Rocky Mountain Energy's Bear Creek uranium operations, it was pointed out that they operated their thickeners at very high wash rates and accepted the resulting lower underflow densities. Both of these properties operated with manual bed level control.

Based on the encouraging signs shown by the two American operations, we decided to accumulate some data on the effect of high wash versus dry tons per hour of feed while running the units on manual control. Operators were instructed to feed a constant volume to the units by keeping the chart recording of the magnetic flow meter constant. The underflow pinch valves were operated in conjunction with one another. Results were so encouraging we decided to make runs at lower wash rates to fill in results from the whole range of operating conditions.

In September, feed was taken from both A and B banks of CCD. Data was obtained for feed rates from 40-70 DTPH. and wash rates of 360-600 IGPM.

4. DATA

Operating data such as flow rates, pulp densities, torques and levels were recorded on the operators log sheet once per hour. Marcy bucket samples were used to determine pulp densities. Every two hours, samples of the feed, underflows, and overflows were collected. The filtrates from these samples were then analysed for U_3O_8 (grams per litre). Daily composites of the No. 2 enviro clear underflow solids were prepared and then screened to determine the amount of minus 200 mesh material. All the operator log sheets are on file at the Process Development Department. Daily summaries for each month are tabulated in Appendix 1.

5. DISCUSSION

5.1. Mechanical Problems

Keeping the enviro clear test program running was a difficult feat. Spare pumps were taken by Construction for the C Bank expansion. The gear box for the No. 1 enviro clear underflow pump was removed by the mill mechanics to replace a broken down unit. Data could only be obtained for running underflow removal via the pinch valves. Most of the mechanical problems associated with the units can be easily rectified.

5.1.1. Rake Drive

The enviro clears were supplied with a Speedy Vickers HAS 7 hydraulic pump with a 10 h.p. hydraulic motor. Support gussets had to be added to keep the hydraulic motor aligned with the stub shaft. At high torques, the pump would bypass oil causing the rakes to stop. The bypass has been reset for 2000 psi by the mill mechanical department. The supplied flexible rubber coupling between the hydraulic motor and the stub shaft has been a continuing source of problems. On the No. 3 enviro clear, the coupling was replaced with a Falk-8 flexible coupling. Since then, no problems have been encountered with this coupling. Two new couplings were bored out for the No. 1 and 2 units.

5.1.2. Feed Seal

Feed to the enviro clear enters horizontally from the top of thickener into the center feed pipe. Then it is directed straight downward into the feed cone. The rakes are attached to the feed cone. There is a seal between the feed pipe and the cone to keep pulp from short circuiting.

5.1.2. Feed Seal (Cont'd.)

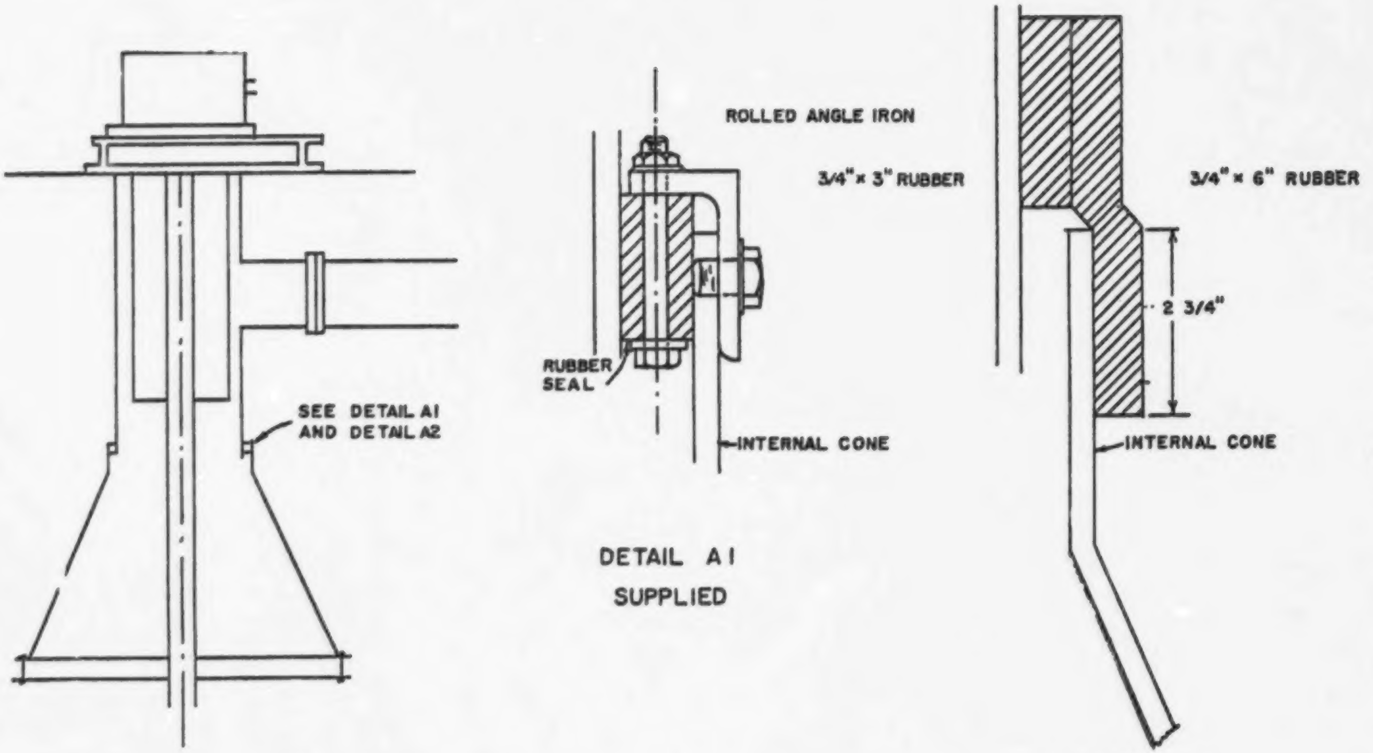
The units came equipped with seals as shown in Figure 6. After operating for a few months, excessive air bubbles were noticed around the feed pipe. Upon shutting down the No. 1 unit, we discovered the rubber seal was gone and its retaining bracket was worn. The feed pipe was also scorred. A new seal, as also shown in Figure 6, was designed and installed on the No. 1 unit. This seal is simply composed of two pieces of linatex rubber attached to the stationary feed pipe. Performance of this seal has been very good.

5.1.3. Feed Distribution

In R & D report R76-02, on the operation of the 30 foot diameter unit, a recommendation was made to install a false cone inside the feed well to stop build-up on the distributor plate. This modification was not carried out on the 35 foot diameter thickeners. The build-up causes problems in evenly distributing the feed. The feed distribution plates are warped. More support will be required to ensure the same gap all around the feed well.

5.1.4. De-aeration Tank

The supplied conical shaped de-aeration tanks are not desirable. Originally, the feed entered tangetially from the side of the cone causing vortexing and engulfing air. Soon after operating the units, an automatic level control system using a pinch valve and a Fisher level-trol float system was installed. The control system worked very well but had to be removed in order to treat high feed rates. To stop the vortexing, the feed lines were repiped to enter the tank vertically from the top. Future de-aeration tanks design should be similar to the flat-bottomed one used for the mine water enviro clear.


SEE DETAIL A1
AND DETAIL A2
ROLLED ANGLE IRON
3/4" × 3" RUBBER
RUBBER SEAL
INTERNAL CONE
DETAIL A1
SUPPLIED
3/4" × 6" RUBBER
2 3/4"
INTERNAL CONE

DETAIL A2
EXISTING

FIGURE 6 FEED SEAL

5.2 Instrumentation

Calibration of all the instrumentation took eight months to complete. Using Quirke instrument technicians was a mistake for their first priority lies with keeping the mill operating. If a project of this magnitude is to be run again, one technician should be assigned to look after the calibration of all the instruments. Most of the instrumentation performed quite satisfactorily.

5.2.1. Dry Tons Per Hour

The magnetic flow meter worked very well. The range on the feed density gama gauge was too narrow. A lot of the time we were feeding pulp at a density of 1150 gm/l. The gamma gauge output drifted a lot. Discussions with Ohmart personnel indicate that our units are not sealed properly. Due to these shortcomings, hourly Marcy density samples of the feed were taken. The operator then calculated the dry tons per hour by using the following equations.

$$\text{FEED RATE} = \text{Mag. Volt} \times 144 \text{ USGPM} \quad \text{(mag. volt 0-10v)}$$

$$\%\text{ solids} = \frac{SG\,(SP - 1)}{SP\,(SG - 1)} \times 100\%$$

SG specific gravity 2.7

SP feed density

f% solids = fractional % solids

$$\text{DTPH} = \frac{\text{USGPM}}{4.02\left(\frac{1 - f\%\text{ solids}}{f\%\text{ solids}} + \frac{1}{SG}\right)}$$

For example:
Mag. flow meter output - 5 volts
Feed density - 1150 g/1
Specific gravity of dry ore - 2.7

Then:

$$\text{Feed rate} = 5 \times 144 = 720 \text{ USGPM}$$

$$\% \text{ solids} = \frac{2.7\ (.15)}{1.15\ (1.7)} \times 100\ \% = 20.72\%$$

$$\text{DTPH} = \frac{720}{4.02\left(\dfrac{1 - .2072}{.2072} + \dfrac{1}{2.7}\right)} = 42.7 \text{ DTPH}$$

The basic instrumentation will work. The span of the gamma gauge will have to be expanded. The existing multiplying board will have to be changed. The drifting of the gamma gauge is currently being investigated by mill personnel.

5.2.2. Underflow Density

The ohmart underflow gamma density gauges output were also subject to drift. Marcy bucket samples were taken along with the gamma gauge readings.

5.2.3. Automatic Flocculant Addition

Several attempts were made at using the automatic flocculant system. The drifting of the gamma gauges cut short our attempts for long-term tests. During our short runs, it was apparent that the underflow density portion of the signal had too much influence. We shortened the density controller's output span from 0-10 to 3-8 volts. A better control approach might be to base the flocculant addition on a lbs per ton basis. The operator could set a certain lbs/ton on a controller and then check a couple of hours later and make whatever adjustments he deemed necessary.

5.2.4. Underflow Removal - Bed Level Control

Testing with underflow removal via the diaphragm pumps was not tried during the test program as the gear box for the No. 1 unit was missing. Over the winter, we had ample opportunities to observe their performance while running with the underflow pumps on manual as well as automatic control. A lot of attention was required to operate the units in conjunction with one another. The best results were obtained when the slow speed stroke of the pump ran at least 75% of the time. Operators had to adjust the pump strokes about once per hour.

Our results using the pinch valve on automatic control were disappointing. No matter what combination of high and low level probe elevation differences and cycle timer OFF and ON increments were used, we still created large surges between the units. Outokumpu has offered us a bed level control system to try on a trial basis. Their system allows for a proportional signal to be sent to the pinch valve. It appears with their system we should be able to operate somewhere between wide open and closed.

Manual operation of the pinch valves in conjunction with one another and keeping a constant volume of feed to the No. 1 unit was our most successful mode of operation. Approximately every thirty minutes to an hour, the operator would check the bed levels and then adjust the air supply pressure to the pinch valves.

5.2.5. Rake Torque

The cut out protection and instrumentation to monitor the rake torque works very well.

5.3. Results

During June and July, we treated approximately the same tonnage throughput as one conventional CCD bank. It was necessary to control the feed rate from the large changes that occur when the CCD underflow pump stroke was changed. At times, feed was taken from A5 and B5 and other times just part of B5. Providing a relatively steady feed tonnage rate (± 10 T.P.H.) was necessary to obtain reliable results.

5.3. Results (Cont'd.)

Under steady conditions, the enviro clears operated as well as the conventional thickeners.

5.3.1. Gap Setting

The distribution plate gap setting was set at 1 5/8 inches. At a nominal tonnage of 60 T.P.H. and 350 IGPM of wash, the pulp exit velocity is approximately 46 fpm. During surges, the bed appeared to be expanded. First, we tried slowing the exit velocity down and hopefully decreasing the bed turbulence by opening the gap on the No. 2 unit to 2 inches. No. 1 unit was kept at 1 5/8" for comparison. The No. 2 unit operated better producing denser underflows and lower slime levels indicating it could handle more feed. Next, we closed the gap on the No. 1 unit to 1-1/4 inches. The No. 1 unit became much more sensitive to upset from surges. From our results, we concluded that the 2 inch gap was the best.

5.3.2. Rake Speed

The hydraulic rake system can be operated from 0-3+ rpm. Enviro clear had recommended setting the rake speed at 1/2 rpm but had also indicated the slower the better. Our units had originally been set at 1/3 rpm. Experimental runs were tried at 1/2, 1/3 and 1/4 rpm. Low underflow densities were obtained at 1/2 rpm. High underflow densities and high torques were obtained at 1/4 rpm. Average densities and torques were obtained at 1/3 rpm. Operating at 1/4 rpm was too close to the edge. Operators had to constantly be wary of the torque. If the rakes stopped for a power bump, the thickener would have to be washed out. Results at 1/3 rpm were not quite as good as at 1/4 rpm, but the units don't require the constant attention and the units can be re-started after a power bump.

5.3.3. Effect of Fineness of Feed

Operation of the new cyclone plant in the Quirke mill is more efficient than before. Very little coarse material is sent to CCD's. The cyclone split is 50:50 rather than the designed 67% to CCD and 33% to cyclone underflow and then to tails. Data obtained earlier on the thirty foot diameter unit indicated that the thirty-five foot diameter enviro clears would have little problem treating 60 T.P.H. at a wash rate of 300 IGPM. These predictions were based on ore that is considerably coarser than the material currently dealt with. Slimes have a compounding negative effect. CCD underflow densities drop, hence more volume is sent to the enviro clears. More flocculant is required to settle the material

5.3.3. Effect of Fineness of Feed (Cont'd.)

causing poorer enviro clear underflow densities. Lastly, the enviro clears capacity is inversely proportional to the amount of slimes in the feed.

5.3.4. Manual Operation

After shutdown, the enviro clears were run by manually controlling underflow withdrawal rate via the pinch valves. First, we investigated the effect of a high wash rate at different tonnage rates. The enviro clears performed very well. Next, we increased the tonnage and kept the same wash to dry tons ratio. The underflow densities dropped but the enviro clears handled up to 70 T.P.H. at a wash rate of 400 IGPM and gave soluble losses of 20%. Operation of the enviro clears on manual control was far superior than running the units on automatic bed level control.

5.3.5. Flocculant Addition

The enviro clear feed rate was too unstable to determine the best flocculant addition point(s) or optimum flocculant dosage. Generally, it was quite obvious that flocculant consumption was proportional to the amount of fines in the ore as well as the dry tons of feed. Our attempts to use the automatic flocculant system indicated it was probably too elaborate for our needs. Automatic flocculant addition based on the amount of dry tons in the feed with provision for the operator to change the dosage is probably all that is required.

5.3.6. Regression Analysis

The inability to control the process variables made it virtually impossible to develop simple graphical relationships. Simple computer programs are available to perform linear multiple variable least square type approximations. Rio Algom's computer facility unfortunately does not have a Basic Compiler. Energy, Mines and Resources, Elliot Lake laboratories permitted us usage of their facilities for a limited number of runs. Only the data collected after shutdown was evaluated as the earlier data was too erratic.

Three statistically significant equations were developed.

1. % LOSS= -129.3 + .37% -200# + .137 DTPH + .086 FEED DEN
   $R^2 = .83$

2. No 2 UF DEN = 1578 - 2.9% - 200# - 2.7 DTPH + .25 FEED DEN
   $R^2 = .83$

3. % LOSS = 10.27 + .4 DTPH - .0174 WASH + .099% - 200#
   - .0143 CCD UF DEN
   $R^2 = .79$

where R^2 = multiple correlation coefficient
a perfect fit is $R^2 = 1$

These equations will provide reliable predictions on the expected performance within the range studied.

On the conventional thickeners, an acceptable CCD design is for a wash ratio of 2.5 : 1 to 2.7 : 1. Where wash ratio is defined as:

$$\text{Wash ratio} = R = \frac{\text{Liquid in Overflow}}{\text{Liquid in Underflow}}$$

The theoretical soluble loss is defined as:

$$\%\ \text{Sol. Loss} = \frac{1}{1 + R_1 + R_1 R_2 + \ldots} \times 100\% \qquad R_i - i^{th}\ \text{stage wash ratio}$$

For our two-stage application, an acceptable soluble loss would be:

$$\%\ \text{Sol. Loss} = \frac{100\%}{1 + 2.5 + (2.5)(2.5)} = \frac{100\%}{9.75} = 10.3\%$$

The best method of defining the capacity of the enviro clears is to accept that they must obtain no more than 10% soluble losses. Then, the regression equations can be applied to see what levels of the variables are desirable.

For example, at 60 T.P.H., to obtain 10% loss,
Assume 80% - 200# , CCD DEN = 1550
Then, if we apply Eq. 1 (most significant)

% LOSS = -129.3 + .37% - 200# + .137 DTPH + .086 FEED DEN
10 = -129.3 + .37(80) + .137(60) + .086 FEED DEN
101.48 = .086 FEED DEN
FEED DEN = 1180

$$\% \text{ solids} = \frac{2.7\ (.18)}{1.7\ (1.18)} \times 100\% = 24.23\%$$

@ 60 T.P.H. and 24.23% solids

Total wt.	= 247.65	TPH
Less solids wt.	60.0	TPH
Total Liq. Wt.	187.65	TPH
Less CCD Liq.	46.5	TPH
WASH LIQ. WT.	141.15	TPH

Then wash volume = 471 IGPM
and the No. 2 E.C. Underflow density is:

No. 2 UF DEN = 1578 - 2.9% - 200# - 2.7 DTPH + .25 FEED DEN
= 1578 - 2.9(80) - 2.7 (60) + .25 (1180)
= 1479

At 50 T.P.H. which is more typical of current CCD operations, the wash rate required to obtain a 10% loss is:

Assume: 80% - 200#, 1550 CCD DEN

Then:

$$10 = -129.3 + .37\% - 200\# + 1137 \text{ DTPH} + .086 \text{ FEED DEN}$$

$$10 = -129.3 + 29.6 + 6.85 + .086 \text{ FEED DEN}$$

$$102.85 = .086 \text{ FEED DEN}$$

$$\text{FEED DEN} = 1196$$

$$\% \text{ solids} = \frac{2.7\ (.196)}{1.7\ (1.196)} \times 100\% = 26.03\%$$

Then, @ 50 TPH and 26.03% solids

Total wt.	=	192.1 TPH
Less solids wt.		50 TPH
Total Liq.Wt.		142.1 TPH
Less CCD Liq.		38.7 TPH
WASH LIQ. WT.		103.4 TPH

Then wash volume = 345 IGPM

And the No. 2 E.C. Underflow density is

$$\begin{aligned} \text{UF No. 2} &= 1578 - 2.9\% - 200\# - 2.7 \text{ DTPH} + .25 \text{ FEED DEN} \\ &= 1578 - 232 - 135 + 299 \\ &= 1510 \end{aligned}$$

These equations are definitely biased in favour of high wash rates. Admittedly, at the high wash rates, the underflow densities drop but the overall uranium loss is lower. Use of these equations on extrapolated data may be totally erroneous. The predicted versus actual % loss figures for August and September are plotted in Figure 7. The trend shows that we have a reliable fit.

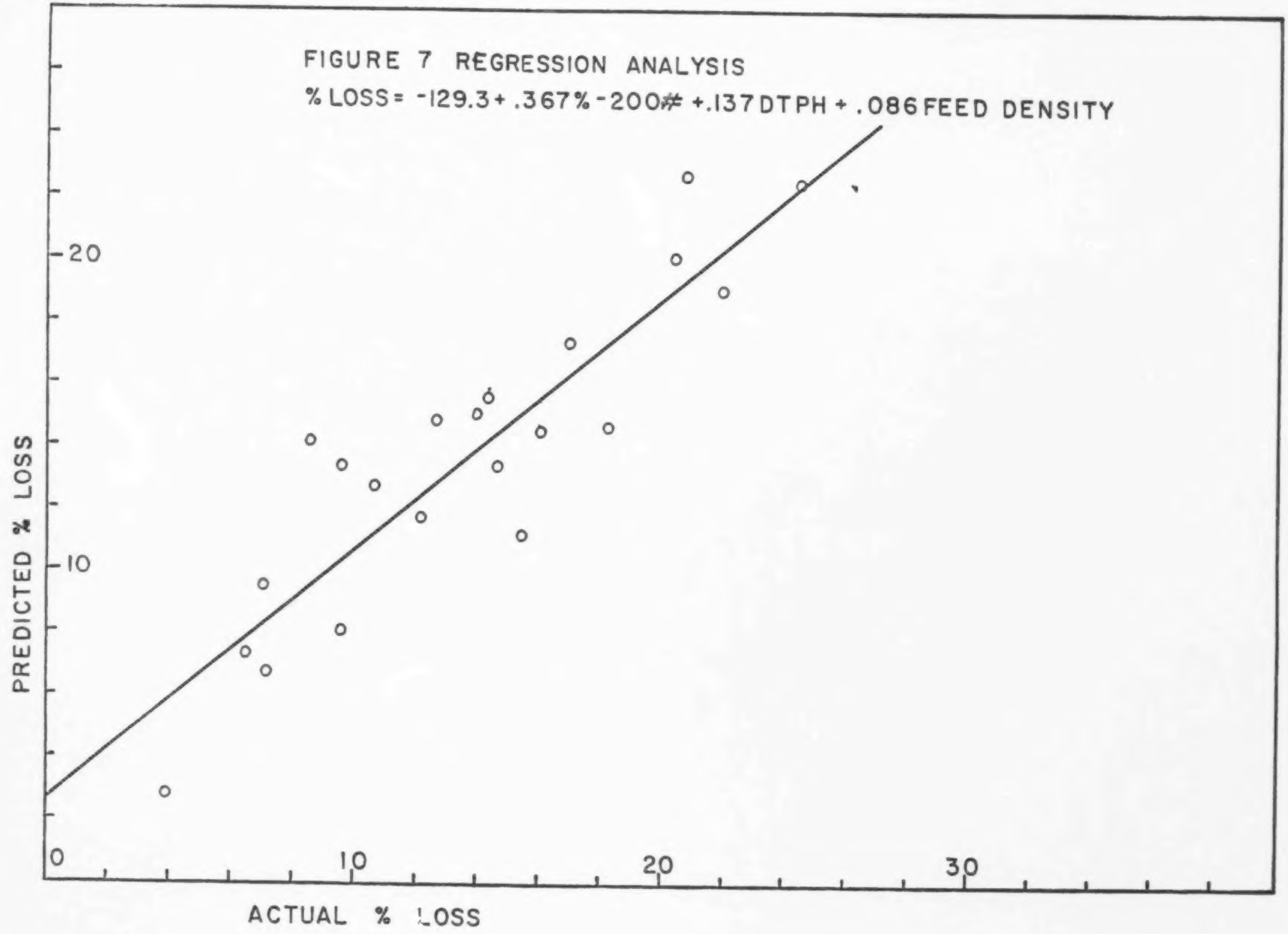


FIGURE 7 REGRESSION ANALYSIS

% LOSS = -129.3 + .367% - 200# + .137 DTPH + .086 FEED DENSITY

6. CONCLUSIONS

The results obtained running on manual control prove that the enviro clears are a viable alternative to the conventional thickeners. The various mechanical difficulties encountered running the test program can be easily overcome. A few modifications of the instrumentation scheme will be required before the units can be fully automated. A less sophisticated automatic flocculant addition system will accomplish the desired results. Control of the underflow withdrawal via the pinch valve should be investigated further. Operation of the units revealed that they are sensitive to surges created within their own circuit. Our best results were obtained by keeping the total feed volume to the No. 1 unit relatively constant while running at high wash rates. This was achieved by making proportionately equal changes to the underflow pinch valves. For example, if the bed level was rising in the No. 1 unit, the operator would open the pinch valve on No. 1 unit as well as on the No. 2 unit. He would keep checking the output from the magnetic flowmeter recorder until it had steadied at the same total flow before any of the changes were originated. Surges from CCD up to 80 DTPH of total feed can be handled as long as they are detected. Sufficient flocculant must be added with the surge. Conditions were never steady enough to optimize the flocculant dosage, or determine the best addition point(s).

Capacity of the units are dependent on many variables. The equations developed using the regression analysis techniques are probably the best way of determining what losses will occur for a given set of conditions.

7. RECOMMENDATIONS

7.1. Mechanical

Future enviro clear thickener installations should incorporate the following mechanical modifications:

1. Support gussets to ensure alignment between the rakes hydraulic drive motor and the gear reducer. Falk drive couplings.
2. Rubber seal between the feed pipe and the rake support. Figure 6.
3. False cone inside the feed well to improve feed distribution and reduce blockages. More support for the distribution plate.
4. Flat-bottomed de-aeration tanks, fed vertically from the top.

7.2 Instrumentation

Most of the instrumentation performs quite well. The rake monitoring and cut off protection saved us on a few occasions from overloading and perhaps damaging the rakes. Not enough time or manpower was available to solve our problems with the output from the gamma gauges drifting. These units should either be repaired or replaced. We obtained little success with the automatic bed level control scheme, using the underflow pinch valve and the ultra sonic probes. This system should be abandoned and a proportional system like the one suggested by Outokumpu Instruments be evaluated. The automatic flocculant addition system seemed too elaborate for our needs. For future applications, a lbs/ton system with provision for the operator to alter the lbs/ton ratio will suffice.

7.3. Operations

To obtain the lowest soluble losses, the enviro clears should be operated at high wash rates. The data and regression equations developed from this test program should be used in determining the capacity and performance for future applications of enviro clear thickeners as CCD units on Elliot Lake ores.

8. ACKNOWLEDGEMENTS

I would like to thank the mill operating mechanical, electrical and instrument personnel who assisted us during the test program. Without their help, we could never have hoped to run the test program. Also, thanks are extended to the staff at Energy, Mines and Resources, Elliot Lake laboratories for usage of their computer facility. Assistance from our University of Waterloo Co-op Students in gathering the data and preparing this report was appreciated. Lastly, thanks are due to my supervisor, Mr. E. Barnes, who proposed I make this presentation.

9. BIBLIOGRAPHY

1. Mather, J.T., Mine Water Treatment in an Enviro-Clear Clarifier, R75-08 (November 1975)

2. Arthur, D.B., Pilot Plant Test of Enviro-Clear Thickener on Neutral Thickener Feed, R75-09 (November 1975)

3. Mather, J.T., Pilot Testing of a 3' Enviro-Clear Thickener for CCD Substitution, R75-10 (December 1975)

4. Arthur, D.B., 30 Foot Diameter Enviro-Clear as Sixth Stage of CCD, R76-02 (March 1976)

5. Draper, N. R., & Smith, H., Applied Regression Analysis, Wiley, 1967

6. Mular, A.L., & Bull, W.R., Mineral Processes: Their Analysis, Optimization and Control, Queens University, 1971

TABLE 1 - JUNE DAILY AVERAGES

DATE	DTPH	WASH IGPM	UNDERFLOW DENS			FEED DEN	FLOCC LB/TON	%-200#	DILUTION		% SOL LOSS	USGPM FT^2	GAP RAKE		
			CCD	NO 1	NO 2				B5/B6	B6/B7			NO 1	NO 2	RPM
June 1	43.8	279	1566	1606	1635	1208	.022	77.2	2.62	2.34	7.3	.55	1-5/8	1-5/8	1/3
June 2	46.9	301	1601	1522	1483	1211	.021	76.5	2.69	2.37	14.5	.58	"	"	1/3
June 5	45.3	318	1567	1502	1494	1196	.021	71.7	2.73	2.36	12.1	.61	"	"	1/2
June 6	37.2	304	1480	1512	1480	1167	.021	86.2	1.83	2.53	9.1	.58	"	"	1/2
June 7	42.5	312	1547	1515	1451	1187	.022	85.3	2.47	1.90	12.8	.59	"	"	1/2
June 8	40.5	287	1545	1485	1460	1192	.027	85.7	2.20	2.07	13.7	.55	"	"	1/4
June 9	47.3	241	1531	1526	1497	1232	.029	83.9	1.81	1.61	17.9	.54	"	"	1/4
June 12	55.2	185	1489	1505	1618	1274	.017	70.7	3.09	4.28	19.1	.53	"	"	1/4
June 13	48.4	281	1509	1566	1571	1215	.023	79.7	2.64	2.41	10.4	.60	"	"	1/4
June 14	43.2	270	1487	1590	1581	1199	.025	71.6	2.25	2.34	8.5	.57	"	"	1/3
June 15	42.8	235	1526	1571	1550	1222	.029	77.7	1.86	2.33	12.3	.51	"	"	1/3
June 16	49.8	257	1650	1697	1650	1249	.014	77.0	2.10	2.23	9.0	.52	"	"	1/3
June 20	49.1	216	1540	1515	1629	1254	.026	82.7	2.06	2.35	14.3	.51	"	"	1/3
June 21	48.1	202	1543	1532	1563	1261	.029	83.1	2.49	2.08	18.4	.49	"	"	1/3
June 22	48.7	232	1556	1594	1616	1246	.029	76.0	2.24	2.39	12.1	.52	"	"	1/3
June 23	51.1	212	1603	1594	1698	1275	.019	78.2	2.19	2.16	12.0	.49	"	"	1/3
June 26	53.2	213	1539	1512	1554	1266	.031	85.2	1.81	2.46	20.5	.53	"	2	1/3
June 27	46.7	208	1548	1566	1580	1254	.035	85.2	1.98	2.47	15.2	.48	"	"	1/3
June 28	55.9	229	1519	1569	1577	1257	.034	84.4	1.58	2.27	16.7	.57	"	"	1/3
June 29	43.6	269	1529	1564	1597	1206	.024	84.4	2.59	1.84	9.0	.55	"	"	1/3

TABLE 1 - JULY DAILY AVERAGES

DATE	DTPH	WASH IGPM	UNDERFLOW DENS			FEED DEN	FLOCC LB/TON	%-200#	DILUTION		% SOL LOSS	$\frac{USGPM}{FT^2}$	GAP		COMMENTS
			CCD	NO 1	NO 2				B5/B6	B6/B7			NO 1	NO 2	
July 3	47.9	216	1451	1478	1469	1229	.044	95.8	1.87	1.55	22.2	.55	1-1/4	2"	Rakes set
July 4	55.2	263	1481	1455	1483	1230	.047	94.0	1.81	1.89	21.2	.63	"	"	at 1/3 RPM
July 5	54.8	266	1532	1488	1506	1238	.041	93.9	1.41	1.81	19.5	.60	"	"	RPM's
July 6	38.8	278	1576	1614	1595	1194	.022	84.7	1.88	1.76	6.9	.51	"	"	
July 7	60.3	244	1651	1684	1653	1289	.017	89.8	3.03	1.39	13.2	.54	2	"	
July 8	51.1	224	1481	1503	1544	1240	.025	82.7	2.30	2.05	17.7	.55	"	"	
July 9	40.3	273	1545	1480	1489	1198	.034	82.2	3.00	3.49	13.3	.56	"	"	
July 10	32.5	255	1544	1431	1521	1180	.050	89.9	4.67	3.14	10.3	.52	"	"	Density
July 11	46.3	218	1458	1410	1426	1226	.044	93.8	1.68	2.06	27.7	.57	"	"	cup plugged
July 12	56.5	206	1475	1487	1455	1260	.047	89.9	1.26	1.84	31.0	.56	"	"	No 1 E.C.
July 13	51.4	199	1515	1460	1610	1263	.045	86.7	1.31	1.91	18.5	.51	"	"	
July 14	62.8	169	1489	1468	1518	1300	.036	78.3	1.25	1.94	34.8	.55	"	"	
July 15	60.8	238	1484	1414	1377	1254	.031	91.0	1.33	1.63	44.5	.63	"	"	
July 16	60.4	245	1499	1428	1395	1254	.040	83.3	1.53	1.95	39.6	.62	"	"	
July 17	59.1	218	1468	1413	1455	1257	.041	84.3	1.20	1.55	32.7	.60	"	"	
July 18	45.6	286	1485	1469	1464	1198	.037	87.1	1.65	1.97	15.5	.60	"	"	
July 19	36.3	301	1481	1522	1590	1166	.026	84.1	1.86	2.15	6.0	.57	"	"	
July 20	46.6	298	1527	1496	1503	1201	.020	82.0	1.76	2.11	13.1	.60	"	"	
July 21	43.0	313	1524	1449	1485	1185	.028	86.2	1.54	1.52	12.5	.61	"	"	

TABLE 1 - AUGUST DAILY AVERAGES

DATE	DTPH	WASH	UNDERFLOW DEN			FEED	FLOCC	%-200#	DILUTION		% SOL	USGPM	COMMENTS
		IGPM	CCD	NO 1	NO 2	DEN	LB/TON		B5/B6	B6/B7	LOSS	FT^2	
Aug. 21	43	404	1533	1575	1640	1170	.014	70.0	3.37	2.50	4	.65	Rakes set at 1/3 RPM
Aug. 22	48.8	478	1578	1528	1501	1157	.017	83.0	2.98	2.85	6.5	.83	
Aug. 23	63.3	488	1506	1416	1418	1184	.025	81.7	3.10	2.41	15.5	.92	Gaps set at 2"
Aug. 24	50.1	514	1491	1450	1405	1143	.022	87.8	2.43	2.08	9.6	.91	
Aug. 25	58.6	504	1511	1399	1397	1180	.019	90.5	2.19	1.86	14.7	.90	Runs at ≈ 20% Solids
Aug. 26	51.9	462	1497	1487	1530	1161	.019	86.9	3.81	2.94	7.1	.86	
Aug. 28	68.1	629	1529	1533	1600	1167	.010*	77.5	4.31	1.94	5.0	1.1	* Over flocced in morning Runs at 30% Solids.
Aug. 29	75	486	1533	1391	1388	1233	.021	95.5	2.26	1.93	24.5	.89	
Aug. 30	72.4	460	1543	1428	1424	1206	.026	97.8	2.34	2.12	20.5	.96	
Aug. 31	49.3	363	1513	1493	1529	1198	.026	89.6	2.41	2.50	9.6	.69	

TABLE 1 - SEPTEMBER DAILY AVERAGES

DATE	DTPH	WASH	UNDERFLOW DEN			FEED	FLOCC	%-200#	DILUTION		% SOL	USGPM	COMMENTS
		IGPM	CCD	NO 1	NO 2	DEN	LB/TON		B5/B6	B6/B7	LOSS	FT^2	
Sept. 1	54.6	365	1543	1471	1484	1200	.025	91.8	1.29	1.14	14.0	.74	Rakes set at 1/3 RPM
Sept. 2	63.8	420	1511	1476	1471	1203	.028	89.0	1.28	1.54	14.4	.86	
Sept. 3	56.7	367	1473	1458	1520	1170	.024	89.2	3.11	1.81	12.2	.90	
Sept. 4	64.5	409	1450	1497	1490	1191	.023	89.7	1.86	2.06	12.7	.89	Gaps set at 2"
Sept. 5	63.9	419	1490	1423	1446	1203	.038	94.0	2.11	2.11	17.0	.84	
Sept. 6	72.9	391	1540	1437	1461	1287	.026	75.5	2.23	2.08	22.0	.84	
Sept. 7	52.8	390	1450	1376	1395	1208	.038	89.5	1.89	1.86	18.3	.81	
Sept. 8	46.9	504	1500	1451	1461	1181	.019	76.5	2.31	2.13	7.2	.85	
Sept.14	71	339	1571	1501	1503	1310	.038	80.8	1.83	1.65	20.8	.73	
Sept.15	66	396	1563	1473	1494	1226	.041	80.0	1.55	1.53	16.1	.84	
Sept.20	62.2	604	1522	1420	1421	1205	.034	81.5	2.40	1.87	10.7	.96	
Sept.22	71.8	600	1615	1533	1504	1220	.027	78.3	2.05	1.68	8.6	1.05	
Sept.28	60.5	521	1514	1504	1443	1201	.042	77.6	2.01	1.63	10.1	.99	

Discussion of Paper by L. Hamel

E. Joe: Where is the main economic advantage of installing Enviroclear thickeners ?

Reply: I believe it is in the capital cost. An Enviroclear costs about a half a million whereas a conventional unit costs a million and a half.

W. Meekel: Whose ultrasonic level indicators do you use ?

Reply: We use Sensol probes handled by Brian Engineering of Toronto.

R. Colqhoun: At what pulp density do you run the thickeners ?

Reply: We normally use polymer flocculants in the C.C.D-circuits. The pulp denisty is around S G 1500. In the conventional unit it is about S G 1600. The Enviroclears produce no adverse effects on ion exchange.

APPENDIX - A

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